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Phase 3 Final Report

Cost Sensitivity Study of Major Process Parameters

Submitted To

DEPARTMENT OF HEALTH, EDUCATION, AND WELFARE
PUBLIC HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

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Cost Sensitivity Study of Major Process Parameters

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DEPARTMENT OF HEALTH, EDUCATION, AND WELFARE
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Contract No. PH 86-68-86

by

THE M. W. KELLOGG COMPANY
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March 31, 1970

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I. INTRODUCTION

The work described in this report was performed under Contract No. PH 86-68-86, Department of Health, Education, and Welfare, Public Health Service, Consumer Protection and Environmental Health Service, National Air Pollution Control Administration (NAPCA). The tasks designated by the subject contract are divided into three phases: phases 1 and 2 covering data evaluation, preliminary process design, optimization and economics, and phase 3, as modified at the end of the first two phases, comprising a cost sensitivity analysis of the major process variables. A separate report was issued (February 1970) for phases 1 and 2; phase 3 work is the subject of this report.

The objective of the phase 3 work was to provide a means for making rapid, yet reasonably accurate, preliminary economic evaluations of dry, regenerable type SO_2 removal processes based on a minimum of experimental data. This will permit not only a means of judging the merits of alternative processes but also will establish the relative importance of processing steps or design considerations. The latter will be of value in establishing the worth and direction of experimental programs.

Bases for achieving this objective are the base case designs and economic evaluations prepared for the alkalized alumina process in the phases 1 and 2 work. In the phase 3 evaluations the major process parameters are varied over a range of values and the effect of each parameter on both plant investment and operating cost is then determined.

Results are reported in the form of charts showing the change in investment and operating costs from the base case values for the various parameters studied. Although ideally it would be desirable to have a separate chart for each variable, the number of variables studied make this approach impractical and therefore combinations have been used in many cases. Furthermore, because of the interaction among certain variables, all of the results reported are not completely additive. That is, the change in investment and operating costs shown on the individual charts is derived from the base case conditions and when results from several different charts are added together they may not show the same effect that would be obtained if the combination itself were evaluated. The discrepancy in the chart results is expected to be small, however, and use of these charts to obtain preliminary investment and operating costs of proposed SO_2 removal processes should permit valid evaluations to be made.

In scope the present study was limited to dry, regenerable type processes treating hot ($\geq$300°F) gases containing low concentrations of sulfur oxides; i.e., power plant flue gases rather than industrial stack gases such as, for example, the effluent from a sulfuric acid plant.

It should be noted also that while the base case costs are for alkalized alumina many of the process parameters studied were expanded in range to include other solid sorbents. In the latter case, of course, a number of assumptions are required as to reaction rates, sorbent capacity, method of regeneration, etc. Hopefully, the range of values selected for each parameter will include the conditions likely to be required by dry sorbent processes. For those cases where the conditions are outside the range reported, it may be possible to extrapolate since the trend frequently is obvious.

For ease of comparison, the base case flow sheets and costs are included as part of this report. Three base cases are included, viz, designs based on fixed bed, fluid bed, and dispersed phase process schemes. All the variables studied in phase 3 have not been evaluated for all three designs since some of the parameters caused very little change in the process that was evaluated and it was obvious that similar results would be obtained for the other processes. If the effect of a specific parameter is desired for one of the designs not evaluated, however, the percentage change obtained for the process that was evaluated probably can be used with very little error.

Although the material presented herein provides a quick means of evaluating potential SO_2 removal processes, the results are preliminary and promising candidates should be evaluated separately. Thus, while these methods are satisfactory for comparative ratings and for determining approximate costs, processes which appear to be borderline (or better) in costs may require further study before valid decisions regarding their merits can be reached. Within these limitations, however, the enclosed material should be quite valuable in guiding future development work in dry SO_2 removal processes.

II. SUMMARY

A cost sensitivity study has been made in which the effects of major process parameters on plant investment and operating cost have been determined. A tabulation of the parameters studied is included. Results are reported as changes from the base case values which were obtained from the previously completed phases 1 and 2 evaluations. Three battery limits base case process designs were developed in phases 1 and 2 (dispersed phase, fluid bed, fixed bed), but not all of the variables were evaluated in phase 3 for three designs. In some cases it was apparent from the results obtained for one of the processes that certain variables have only minor effects on investment and operating costs and therefore extending the study to the other two processes was not warranted.

A copy of the flow sheet, process description, battery limit investment and operating cost for each of the base case designs is included for reference. Results are given in graphical form such that changes in the base case values are shown as a function of the different variables studied in phase 3; these changes can be added to the base case values to obtain the investment and operating cost for any specific set of conditions desired. Although some of the results are not completely additive, the probable error is small and for practical purposes the effects obtained from the various charts can be added together.

A brief summary of the results on economics of the major process variables studied is shown in Table 1. A discussion of the individual items follows.

As a reference point, the base case investment and operating costs also are shown in Table 1; the results of the various evaluations are shown as the percentage change from these base case values.

TABLE 1

EFFECT OF MAJOR PROCESS VARIABLES ON INVESTMENT AND OPERATING COST

PROCESS	[1]CAPITAL INVESTMENT		[2]UNIT COST	
	M $	% CHANGE FROM BASE CASE	$/TON COAL	% CHANGE FROM BASE CASE
BASE CASES				
[3]600/1400				
DISPERSED PHASE	30,720	0	4.53	0
FLUID BED	35,900	0	4.98	0
FIXED BED	40,210	0	5.52	0
[3]300/1400				
FULL SIZE, D.P.[4]	23,130	-24.7	3.72	-17.9
FULL SIZE, FL.B.[4]	25,690	-28.4	3.75	-24.7
HALF SIZE, D.P.[4]	16,450	-46.5	2.68	-40.7
HALF SIZE, FL. B.[4]	17,840	-50.3	2.64	-47.0
[3]300/600				
FULL SIZE, D.P.[4]	22,270	-27.5	3.40	-24.9
FULL SIZE, FL.B.[4]	25,130	-30.0	3.50	-29.7
600° ISOTHERMAL OPERATION				
FULL SIZE, D.P.[4]	28,790	- 6.3	4.16	- 8.2
FULL SIZE, FL.B.[4]	34,950	- 2.7	4.78	- 4.0
FULL SIZE, FX.B.[4]	30,550	-24.0	4.26	-22.8
1400° THERMAL REG.				
DISPERSED PHASE	28,890	- 6.0	4.01	-11.5
FLUID BED	34,860	- 2.9	4.54	- 8.7
1.5% SULFUR IN COAL				
DISPERSED PHASE	26,050	-15.2	3.74	-17.4
FLUID BED	31,630	-11.9	4.52	- 9.2
6% SULFUR IN COAL				
DISPERSED PHASE	39,650	+29.1	6.04	+33.3
FLUID BED	40,960	+14.1	5.51	+10.6
500 MW PLANT				
DISPERSED PHASE	20,510	-33.2	5.95	+31.3
FLUID BED	23,610	-34.2	6.56	+31.9
1500 MW PLANT				
DISPERSED PHASE	39,940	+30.0	3.97	-12.4
FLUID BED	46,280	+28.9	4.29	-13.9

[1] 1000 MW POWER PLANT INVESTMENT IS ABOUT $140-$180 MILLION

[2] POWER PLANT FUEL COST = $6.50/TON

[3] INDICATES SORPTION & REGENERATION TEMPERATURES, °F

[4] D.P.=DISPERSED PHASE, FL.B.=FLUID BED, FX.B.=FIXED BED

1. PROCESS VARIABLES

A. SORPTION STUDIES

- Maintaining a regeneration temperature of 1400°F while reducing the nominal sorption temperature from the base case value of 600°F to 300°F (i.e., the flue gas is treated downstream of the air preheater instead of upstream) provides a more economical design than does the base case. Investment savings of about 25% are achieved while operating costs are reduced by 18-25%. Of particular interest is the fact that at the lower sorption temperature the difference in unit costs between the dispersed phase and fluid bed essentially disappears. The fixed bed was not evaluated for these conditions since no large savings were apparent; i.e., the sorbent inventory would remain the same since the total amount of sulfur removed does not change. Thus the fixed bed sorbers would not be reduced in size and the only savings would be in pressure drop owing to the lower volume of flue gas at 300°F.

It should be noted that while only a few sorption data are available at 300°F, it appears that operation at this temperature is feasible for some of the solid sorbents and thus the potential savings indicated are felt to be valid.

- The above discussion applies to a full size SO_2-removal plant; i.e., one sized to treat all of the flue gas emitted from a power plant during peak power generation. Another scheme has been evaluated whereby the SO_2-removal plant is sized to treat all the power plant flue gas during off-peak power generation or about one half of the peak flow. During peak power demand periods it is assumed that low sulfur coal will be burned in half of the power plant boilers while treating the flue gases from the other half which will be burning high sulfur coal.

This latter design not only permits the SO_2-removal plant investment to be greatly reduced but also results in a higher load factor thereby effecting even further reductions. The net result of using a half-size SO_2 removal plant is a reduction in investment and operating cost of about 50% for both the dispersed phase and fluid bed designs. The respective unit costs for the half-size plant are \$2.68/ton and \$2.64/ton not including any additional costs that might be incurred owing to the need for storing, handling, and burning both low and high sulfur coal in the same power plant. Nevertheless the potential savings involved indicate that this mode of operation should be considered for specific plants before final selection of a SO_2-removal plant is made.

- A half-size fixed bed scheme was not evaluated since it did not offer any large savings. That is, while one train of equipment could be eliminated for the dispersed phase and fluid bed, the fixed bed design would retain the same, albeit smaller, equipment and consequently the savings would be much smaller. This of course does not rule out a half-size fixed bed for some specific location and/or particular set of conditions but these would require an individual evaluation which is beyond the present scope of study.

B. REGENERATION STUDIES

- Evaluation of the economic effect of a 600°F regeneration was made for two different concepts; viz, in combination with sorption temperatures of 300°F and 600°F, the latter of course representing isothermal operation.

The lower regeneration temperature produces only modest savings for the dispersed phase and fluid bed processes, amounting to less than 10% and 5% respectively, as shown in Table 1 under isothermal operation. The effect of lowering the regeneration temperature can be seen also by comparing the 300°/1400°

and 300°/600° cases in Table 1. These latter results show about the same reduction in costs at the lower temperature as is shown by the isothermal operation at 600°F for the dispersed phase and fluid bed designs.

- Isothermal operation at 600°F produces significant reductions in the fixed bed economics, viz, 24% in investment and 23% in unit costs. In fact, the fixed bed investment decreases to the extent that it is lower than the fluid bed and the operating costs approach those of the dispersed phase design. The larger savings for the fixed bed process result from deletion of more equipment and significant reductions in material of construction costs owing to the higher allowable metal stresses obtained at the lower temperature.

It should be noted that no 600°F regeneration data were available at the time of this evaluation. Therefore design parameters (e.g., kinetics, degree of regeneration, etc.) were assumed to be the same as at 1400°F. Thus the indicated savings at the lower regeneration temperature are contingent upon confirmation of assumptions with experimental data.

- Thermal regeneration was evaluated for the dispersed phase and fluid bed processes only based on assumed design parameters for some unspecified metal oxide (e.g., copper oxide). It was assumed that the sorbent is regenerated by heating it to 1400°F in the presence of air such that the sorbed sulfur is removed as SO_2 which is then converted to concentrated sulfuric acid as a salable by-product. Although H_2SO_4 would require more expensive handling and storage facilities than elemental sulfur, no additional costs were included for these items. For the conditions described, the dispersed phase investment and operating costs are reduced by 6% and 11.5% while the corresponding values for the fluid bed are 2.9% and 8.7%.

Fixed bed thermal regeneration was not evaluated since the other two cases showed only modest savings at best and the higher regeneration temperature (1400°F vs. 1200° fixed bed base case) would tend to increase the fixed bed costs, thus reducing the net savings.

C. SULFUR STUDIES

- The economic effect of sulfur concentration in the power plant coal was determined over a 4-fold range, viz, 1.5% S to 6% S. The base case designs and evaluations are for a coal containing 3% S.

The investment and operating costs of the SO_2-removal plant were found to increase with increasing sulfur concentration in the coal but, as is shown in Table 1, not in direct proportion. For example, a 100% increase in the coal sulfur content (from 3 to 6%) causes the dispersed phase investment and operating costs to rise by about 29% and 33% respectively. The costs increase with sulfur content because there is a net loss on the sulfur produced and therefore the greater the quantity made, the higher the loss.

- To determine the potential for making a profit from the by-product sulfur, a breakeven price has been calculated for the various process conditions studied. (Note that the breakeven price is defined as the price at which sulfur must be sold to recover all the costs attributable to the SO_2-removal plant.) The lowest breakeven price obtained is about $90/short ton ($/ST) and is for a fluid bed process sized to handle the maximum flow of flue gas from a 1500 MW power plant burning a 6% sulfur coal. The breakeven prices for the half-size plants were only slightly less than for the full size plants.

- The lowest selling price of the sulfur necessary to recover SO_2-removal plant costs is so much greater than the estimated price (~\$95/ST vs. ~\$20/ST) that it appears essentially impossible to achieve a self-supporting process. To determine the conditions of the SO_2-removal plant itself a second break-even price was calculated based on the reformer and Claus plant only. The reasoning behind this second approach was that a reducing gas and Claus plant always would be required for the production of sulfur. The results of this latter study showed that sulfur must be sold at about \$37/ST based on the dispersed phase and fluid bed processes, and about \$54/ST for the fixed bed scheme. The higher fixed bed cost is owing mainly to the more costly Claus plant pretreatment section needed because of the low pressure of the regenerator effluent gas. Since the cited costs are about the minimum that can be achieved, even with no charges for the SO_2-removal plant, and because they exceed the probable maximum selling price of by-product sulfur, it appears that, at least in the foreseeable future, any SO_2-removal process requiring reducing gas and a Claus plant will always show a net loss. Further the loss will increase with increasing sulfur content in the coal.

- Sulfur content of the loaded sorbent was studied for the dispersed phase and fluid bed designs. Since the nature of the fixed bed operation is such that the sorbent is essentially saturated, a study of sorbent loading was not required for this scheme. Except for the dispersed phase design in which the sorbent loading is reduced to half the base case (i.e., 1-1/2% S), sorbent loading has little effect on economics. The cited option shows increases of 12.5% in investment and 31.3% in operating costs from which it is concluded that a dispersed phase process should be designed for a minimum sorbent loading of about 3%. The results of the sorbent loading study are shown in the body of the report but not in Table 1.

D. PLANT SIZE

In addition to the base case of 1000 MW, power plant sizes of 500 MW and 1500 MW were evaluated for their effect on SO_2-

removal plant costs. It was found that lowering the plant size to 500 MW increases the unit costs by about 31-32% whereas a 12-14% decrease is realized for the 1500 MW size. Extrapolation of the results indicates that very little additional savings would be obtained by increasing the plant size above 1500 MW. On the other hand, indications are that costs will rise rapidly as plant size falls below 500 MW.

The cited costs are for the dispersed phase and fluid bed designs but it is expected that fixed bed costs would show about the same response.

E. OTHER PROCESS VARIABLES

Additional process variables which were studied, such as reducing gas utilization and cost, and degree of SO_2 removal from the flue gas, had relatively minor effects on plant costs. The results of these studies are included in the body of this report but are not shown in Table 1.

2. NON-PROCESS VARIABLES

All of the preceeding discussion applies to what has been defined as process variables as opposed to non-process variables such as capital charge rate, attrition losses, load factor, etc. These latter variables also have been studied to determine their effect on plant costs. Since the degree to which they affect costs depends largely on the values selected, the main body of this report should be consulted for the combinations desired. Only a few selected examples will be discussed here.

A. INVESTMENT, CAPITAL CHARGE RATE, LOAD FACTOR

These items were varied over the following ranges: investment -- $10-$50 million; capital charge rate -- 10-20% of investment; load factor -- 50%-90%, referring to the percentage of time the power plant is operating at full capacity. The effect on plant investment and operating costs for each of the above variables was calculated for the range indicated. From the results of these studies it was found that even with a high load factor (e.g., 80%) and low capital charge rate (e.g., 13%) an investment of less than $20 million is required to reduce the unit cost to about $2/ton coal, which has been mentioned as a reasonable goal.

- Other combinations of the above variables can be evaluated as desired since the results are presented in the form of charts. A word of caution is in order regarding the use of these charts, however, since certain combinations may not represent realistic conditions. For example, if a low capital charge rate is selected, indicating a long plant life, then the load factor should be low also since load factor generally decreases with plant life. Thus in the example cited above the combination of a 13% capital charge rate and 80% load factor may not be obtainable in actual practice. If for a 13% rate, which corresponds roughly to a 30-35 year plant life, the load factor is actually 50%, then to obtain a $2/ton unit cost the maximum allowable investment would be about $10 million. The combination of charge rate and load factor applicable to specific sites will have to be determined on an individual basis using the appropriate charts.

B. ATTRITION LOSSES

- Sorbent losses due to attrition can be a significant cost item, particularly for the dispersed phase design, depending on the attrition rate and unit price of the sorbent. The economic effect of attrition losses have been determined for attrition rates of 0-1% (or in the fixed bed case, a sorbent life of 0-8 years) and unit prices of 5 to 50¢/lb and the results are presented in the form of charts in the body of this report.

At the base case assumed value of 25¢/lb dispersed phase sorbent costs increase by about \$0.50/ton coal per each 0.1% rise in attrition rate. Thus if the actual rate were 0.5% instead of the assumed 0.1% (basis solids throughput rate), unit costs increase by about \$2/ton coal.

- While sorbent losses are less significant in the fluid bed, owing to the lower solids throughput, an attrition rate of 0.5% would add about \$1.25/ton to the base case unit cost.

- Sorbent cost is a function of sorbent life in the fixed bed design and cost studies show that above a 2-year life there is very little effect on unit cost. For example with 25¢/lb sorbent increasing the life from 2 to 8 years reduces the unit costs only by about \$0.10/ton. Below a life of 1 year, however, the costs start increasing rapidly and it is doubtful if a sorbent life of less than about 6 months would be acceptable for a commercial plant.

- It should be emphasized that the range of attrition rates used in the above examples have not been achieved in actual (i.e., pilot plant) operation. Because of the potentially large effect of sorbent losses on costs, therefore, it appears that an attrition rate of well under 1% (or fixed bed life of 1-2 years) must be demonstrated before a sorbent can be considered to be commercially acceptable.

C. MAINTENANCE COSTS

- The effect of plant maintenance charges on operating costs has been determined as a function of plant investment and load factor for the following ranges: maintenance rates -- 2-5% of investment; investment up to \$50 million; load factor -- 50-90%. Results are presented in the form of charts which permit any combination of the three variables to be evaluated for their effect on SO_2-removal plant operating costs.

- The base case annual maintenance charge of 5% of investment was selected based on experience with chemical plants and refineries, but this could be different for the SO_2-removal process. Rates greater than 5% are not included in the present study since the base case costs are already very high and adding additional charges stemming from increased maintenance costs seems superfluous. If the charges are lowered to 3% savings of $0.50 to $1.0/ton over the base case costs are indicated depending on which load factor is used.

3. GENERAL

An example of the use of all the charts is included in the body of the report. It is believed the information presented in these charts can be used to determine approximate economics of potential SO_2-removal processes based on a minimum of experimental data, thereby providing an early process screening method which should be helpful in guiding future research programs.

III. CONCLUSIONS

The following conclusions are based on the results of the study described in this report.

- It presently appears that all processes designed to remove SO_2 from power plant flue gas will increase the cost of power generation. The most economical process therefore is the one which produces the smallest increase in power costs. Selection of the best process for any specific location will require a comparison of the various processes available, including both regenerative and non-regenerative ones, and since this type of evaluation is beyond the scope of the present study, a firm conclusion regarding the relative economic status of the metal oxide process cannot be made at this time.

- A sulfur oxides-removal plant sized to treat only one half of the power plant flue gases shows the lowest cost of all combinations studied and will add about \$2.50/ton to the coal cost. This mode of operation probably will require the burning of low sulfur coal in half of the plant during periods of peak power generation when the power plant is operating at full capacity. During times of off-peak power generation, approximately half the time, only high sulfur coal will be burned since all of the flue gas can be treated during these periods. It is not clear if this scheme is feasible since there are no air quality standards available upon which to base the permissable level of SO_2 emissions, and the practicality of periodically alternating high and low sulfur coal in the same boiler has not been established. Nevertheless, the potential savings dictate a closer look at this concept for any location where a SO_2-removal plant is considered.

- Because a chemically regenerative sorbent will require a reducing gas and a Claus plant (or sulfuric acid plant), and since the cost of sulfur based on these two items alone is greater than the expected selling price, it is concluded that SO_2-removal processes will always cost money and increase the cost of power.

- Since the by-product sulfur is sold at a net loss, the more sulfur produced, the greater the loss. Thus it is concluded that even with an SO_2-removal plant, the most economical operation is one which uses the lowest sulfur coal available at regular prices.

- Sorbent attrition losses, hence cost, must be maintained at a low level if commercially acceptable operating costs are to be achieved. Losses well below 1%, basis solids throughput, are indicated, and, in fact, should approach 0.1%. Sorbent development work, therefore, should be directed toward producing a highly attrition resistant sorbent rather than one with high sorption capacity, if both of these requirements cannot be met.

- If fixed bed operation is anticipated, the regeneration temperature should be as low as practical, (i.e., in the 600-800°F range) to keep costs down. Regeneration temperature is of less importance in the dispersed phase or fluid bed designs.

IV. RECOMMENDATIONS

Based on the results of the phase 3 study, the following recommendations are made:

- Only those sorbents which exhibit adequate attrition resistance for commercial application should be used in any experimental program whose objective is to develop a SO_2-removal process.

- The charts presented in this report should be used to determine approximate costs as soon as possible for any potential SO_2-removal process to which they apply. This procedure will permit those candidates with exorbitantly high costs to be eliminated early in the testing program.

- The installation of a half-size rather than full-size SO_2-removal plant should be considered for any site where a sulfur control process is to be installed.

- Whenever possible, SO_2 sorption data should be obtained at the gas temperatures encountered downstream of the air preheater, viz, about 300°F. The lower temperature offers significant cost reductions compared to the 600°F processes.

V. VARIABLES STUDIED

A. INTRODUCTION

The phase 3 work can be divided into two general categories: 1) those variables which are non-process dependent, and 2) the process dependent parameters. Evaluation of the non-process variables requires a large number of repetitive calculations which can be done rapidly and economically by a computer. Consequently a computer program was developed to perform these calculations. The program was designed so that economic factors (i.e., non-process variables) such as fixed capital investment, sulfur credit, attrition rate, sorbent cost, etc., can be varied over the desired range and the resulting effect on process economics calculated both as annual (dollars/yr) and unit costs (dollars per ton of power plant coal burned). In addition, the program was designed to handle all three of the process schemes developed in phases 1 and 2, viz., fixed bed, fluid bed, and dispersed phase systems.

The calculations involving economic parameters are straightforward, of course, but the changes in investment resulting from process modifications have been prorated from the base case estimate. Although this proration method is not as accurate as making explicit estimates for each case, the large number of changes to be studied make it impractical, both time- and money-wise, to follow the latter procedure. The investments obtained by prorating are, however, sufficiently accurate to provide valid comparisons and conclusions to be made.

All of the phase 3 economic evaluations are derived from the base case econimics developed for each process scheme in phase 1 and 2. The results of the various evaluations in some cases are presented in the form of graphs, when a range of values has been studied, or as tabulated data when only one value of a variable is used. A list of the variables for phase 3 study is shown in Table 2 and a brief discussion of these is presented in the following sections.

B. SORPTION/REGENERATION

1. REGENERATION RATE

The regeneration rate of the loaded sorbent was varied over a 4-fold range relative to the base case value. Specifically, rates of one half and twice the standard rate were evaluated.

2. SORPTION/REGENERATION TEMPERATURES

The sorption temperature used in the base case is 600°F, corresponding to the flue gas temperature upstream of the boiler air preheater. Sorption was evaluated also at 300°F, which is the temperature of the flue gas just downstream of the air preheater.

The base case regeneration temperature is 1400°F. The effect of lowering the regeneration was desired and a 600°F level was selected for study since this corresponds to an isothermal operation when sorption is upstream of the air preheater.

3. SO_2-REMOVAL EFFICIENCY

It was assumed that 90% of the SO_2 in the entering flue gas must be removed in the base case SO_2-removal plant. However, since air quality standards are not clearly defined, and indeed will vary in different locations, the extent of SO_2-removal required may not be 90%. Accordingly, two additional levels of SO_2-removal were selected and evaluated, viz, 70% and 95%.

TABLE 2

PHASE 3 VARIABLES

1. Sorbent attrition rate
2. Sorbent unit cost
3. Sorbent loading, i.e. sulfur pickup per unit weight of makeup sorbent
4. Flue gas sulfur removal efficiency
5. Relative sorption rate
6. Sulfur content of coal
7. Stream efficiency
8. Effect of flue gas volume due to different heating value of coal and power plant heat rate
9. Sulfur credit
10. Capital investment
11. Plant life
12. Sorption temperature
13. Regeneration temperature
14. Relative regeneration rate
15. Regeneration gas utilization
16. Cost of regeneration gas
17. Power plant size
18. Sulfur oxides removal process size
19. Cost of money
20. Sorbent particle size

4. THERMAL REGENERATION

The alkalized alumina sorbent must be chemically regenerated but other sorbents may be developed which can be thermally regenerated. Thermal regeneration at 1400°F was evaluated making whatever process assumptions were needed, since no data were available.

C. PLANT INVESTMENT FACTORS

1. INVESTMENT

Future developments of the alkalized alumina process might show substantially different plant investments and new sorption processes having different investment costs may be developed. Consequently, the economic effects of plant investment were evaluated over the range of $10,000,000 to $50,000,000, which should cover the range for SO_2-removal processes.

2. LOAD FACTOR

The percentage of time that a power plant operates at full load varies with a number of factors such as thermal efficiency, investment, etc. A range of load factors from 50% to 90% was selected for evaluation since this probably will cover most situations.

3. CAPITAL CHARGE RATE

The cost of capital investment in power plants depends on a number of variables and varies among different locations. Since a study of power plant costing procedures was beyond the scope of this evaluation, the capital charge rage was arbitrarily varied from 10 to 20%, which should cover practically all cases.

4. MAINTENANCE COSTS

Maintenance costs are difficult to establish for a first-of-a-kind plant such as required for the SO_2-removal processes. A base case rate of 5% of fixed capital investment per year was selected based on experience with chemical plants and refineries, but additional maintenance rates were also evaluated, viz, 2, 3, and 4%.

D. UTILITIES AND CREDITS

1. NATURAL GAS COST

A natural gas price of 27¢/MM Btu was used in calculating the base case costs. Since this value is the average power plant gas price (1967) it represents what is probably the "summer price" while a year-round price would be higher. Consequently, gas prices up to 70¢/MM Btu were selected for evaluation.

2. STEAM CREDIT

Excess steam is produced by the SO_2-removal plant and it is assumed that the steam can be sold to the power plant and the revenue used to reduce operating costs. The value of the excess steam will vary and therefore two price levels have been used, viz, 25 and 50¢/M lbs. The lower value is equivalent to the fuel cost only while the higher value is often used as a typical steam price.

3. SULFUR CREDIT

Four different values have been used for by-product sulfur selling price, viz, 0, 20, 30, and $40/ton. A base case price of $20/ton was selected as the maximum that probably can be obtained in the near future and it is unlikely that the sulfur selling price will fall outside the indicated range for many years.

E. POWER PLANT OPERATION

1. SULFUR CONTENT OF COAL

The base case SO_2-removal plants are designed to treat flue gases obtained by burning coal containing 3% sulfur. Two additional levels were studied, 1-1/2 and 6% sulfur, to determine their effect on coal sulfur on process economics. It is expected that most of the coal burned will fall within the range selected.

2. HEATING VALUE OF COAL

Most of the coal presently burned in power plants has heating values between 10,000 and 13,000 Btu/lb. These two values were used in the present evaluations -- 13,000 Btu/lb for the base case and 10,000 Btu/lb as an alternative. Flue gas volume can be a function of the coal heating value and hence would affect the SO_2-removal plant design.

3. POWER PLANT HEAT RATE

A heat rate of 8600 Btu/kwh was obtained from the literature and used for the base case designs, this value being the projected rate for future large size power plants. Since the heat rate as well as the heating value of the coal affects the volume of flue gas produced, a higher heat rate was evaluated also, viz, 9000 Btu/kwh. It is presently expected that nearly all of future coal-fired power plants will have heat rates falling within this range, i.e., 8600-9000 Btu/kwh.

4. LOAD FACTOR

Power plants presently have load factors of 55 - 60%; i.e., the power generated is equivalent to that obtained by operating the power plant at full capacity for 55 - 60% of the time. Since load factor can vary substantially depending on local conditions, load factors from 50 to 90% were studied.

F. PLANT SIZE

1. POWER PLANT SIZE

The basic power plant size studied is 1000 MW which was selected as being typical of large size plants. To determine the economic effect of size, power plants of 500 and 1500 MW were evaluated. Results of the studies can be extrapolated to other sizes if desired.

2. SO_2-REMOVAL PLANT SIZE

It has been assumed that the SO_2-removal plant should be designed to treat the maximum flow of flue gas from the power plant even though this rate exists only about half the time. As an alternative, a half-size SO_2-removal plant has been evaluated; i.e., all the flue gas can be treated only during those periods when the power plant is at about half-load. It is assumed that air quality standards can be met during peak loads by burning low sulfur coal in half of the power plant while treating the gases from the other half.

G. SORBENT

1. SORBENT LOADING

The concentration of sulfur on the sorbent leaving the reactor is a function of the process design and operating conditions. The sulfur loading has been varied over a range of values for the fluid bed and dispersed phase designs, viz, 1-1/2 - 6% and 5 - 9% respectively. The nature of the fixed bed operation is such that the sorbent is loaded to essentially 100% of its theoretical capacity and therefore loading is not a variable.

2. SORBENT PRICE

Since a commercial sorbent is not presently available the price cannot be fixed. Estimates indicate that 25¢/lb is a reasonable base point but prices from 5 to 50¢/lb also have been evaluated since the final price probably will fall somewhere within this range.

3. ATTRITION RATE

Attrition rates used in the base case economics are of necessity assumed values since no experimental data are available for commercially acceptable sorbents. Attrition losses of up to 1% of the circulating solids rate have been evaluated for the fluid bed and dispersed phase designs. The sorbent life in the fixed bed process has been varied from 6 months to about 8 years, which should be a sufficiently broad range to cover almost any sorbent that might be used.

H. REGENERATION GAS PRODUCTION

1. REFORMER DESIGN

Two different reformer designs were evaluated - one which produces an effluent gas containing a high percentage of water and one whose outlet gas has less than 10% water. The first design requires a post-reforming treatment to dry the gas whereas the second scheme produces gas that can be used directly.

2. REGENERATION GAS UTILIZATION

The comsumption of reducing gas in the sorbent regenerator was varied from 70% to 90% of theoretical, 80% being the base case. Sufficient experimental data were not available to define the gas consumption exactly.

3. NATURAL GAS COST

Natural gas (methane) is used both as the feed to the reformer and as process fuel. The cost of natural gas was varied from 27 to 70¢/MM Btu to coincide with the range used in the utilities study (Section D.1. above).

4. LOAD FACTOR

The range of load factors studied under power plant operation (i.e., 50 - 90%) also was used in evaluating the cost of regeneration gas production.

5. PLANT SIZE

Reformer plant size was varied over a small range (about ± 10%) to correspond to the 70 - 90% reducing gas utilization studies, and over a wide range (± 50%) to match the power plant size studies.

I. MISCELLANEOUS

Two variables (i.e., relative sorption rate, sorbent particle size) originally selected for study were in fact not evaluated; the reasons for deleting these items are:

A qualitative analysis of the effect of relative sorption rate on process economics revealed only small changes could be expected over the range to be evaluated. For example, in the fluid bed design the reactor size is not set by sorption rate but by mechanical and operational requirements. In the dispersed phase the reactor would change somewhat but economic effects would be small.

- To evaluate the effect of particle size on process design and economics requires experimental data which were not available. Further, it does not appear that small changes in particle size will appreciably affect plant design or economics; large changes in size may have an effect but are indeterminate without data.

- Time and funds were limited and it appeared more beneficial to concentrate on other variables which have greater effects on process design and economics.

VI. EFFECT OF VARIABLES ON PROCESS ECONOMICS

A. INTRODUCTION

In this section, information is presented which shows the effect on plant economics of the variables previously discussed in Section V. Changes in both plant investment and unit costs, relative to the base cases, are shown as a function of process variables and economic parameters. Plant investment refers to a battery limits plant, i.e., offsites are not included. Changes in unit cost are listed as dollars per ton of coal burned in the power plant.

Where possible, i.e., where the variables are numerically continuous, the cost data are presented in graphical form. In other cases, where the effect of the variable cannot be illustrated graphically, the cost data are tabulated. An example of this latter kind of variable is the type of sorbent regeneration employed, chemical or thermal.

B. SORPTION/REGENERATION

1. REGENERATION RATE

The effects of changes in the sorbent regeneration rate on plant investment and unit costs have been determined, and the results are shown in Tables 3 and 4. For each design, the regeneration rate has been varied from one-half to twice the base case rate. The regeneration temperature has been kept constant, at 1400°F, for all cases.

A 30 minute holdup was used as a design basis for the fluid bed case in order to reduce the sulfur content of the sorbent from 7.0 to 2.6 wt %. For the minimum and maximum rates then holdup times of 15 minutes and 1 hour respectively are required.

TABLE 3

EFFECT OF REGENERATION RATE ON PLANT INVESTMENT

DISPERSED PHASE

RATE RATIO	0.5	1.0	2.0
CHANGE IN INVESTMENT ($)	+526,600	-	-306,400
PLANT INVESTMENT ($)	31,250,000[1]	30,720,000[1]	30,410,000[1]
% CHANGE IN INVESTMENT	+1.7	-	-1.0

FLUID BED

RATE RATIO	0.5	1.0	2.0
CHANGE IN INVESTMENT ($)	+139,900	-	-83,400
PLANT INVESTMENT ($)	36,040,000[1]	35,900,000[1]	35,820,000[1]
% CHANGE IN INVESTMENT	+0.4	-	-0.2

[1] ROUNDED TO NEAREST $10,000

TABLE 4

EFFECT OF REGENERATION RATE ON OPERATING COST

DISPERSED PHASE

RATE RATIO	0.5	1.0	2.0
CHANGE IN UNIT COST ($/TON COAL)	+0.06	-	-0.04
UNIT COST ($/TON COAL)	4.59	4.53	4.49
% CHANGE IN UNIT COST	+1.3	-	-0.9

FLUID BED

RATE RATIO	0.5	1.0	2.0
CHANGE IN UNIT COST ($/TON COAL)	+0.02	-	-0.01
UNIT COST ($/TON COAL)	5.00	4.98	4.97
% CHANGE IN UNIT COST	+0.4	-	-0.2

In the dispersed phase design, a moving bed regenerator is employed. The design criteria used for this vessel in the base case are:

1) decrease in sulfur content from 3.0 wt % to 0.5 wt %;
2) a gas velocity of 0.3 feet per second or less, to avoid fluidization;
3) a two hour residence time at normal conditions;
4) an allowance in vessel height to provide a 50% safety factor on residence time.

For the minimum and maximum rates, then, holdup times of 1 hour and 4 hours are obtained, respectively.

As can be seen from Table 3 the effect of changing the regeneration rate, hence solids residence time, is small when compared to the total investment. In the moving bed case, the increase in investment for the minimum rate is $526,600 of which $353,000 is due to the increased sorbent holdup. In the fluid bed case at minimum rate, the increase in investment is $139,900 of which $40,000 is due to increased sorbent holdup. Since the sorbent holdup time is directly proportional to the reaction rate, the holdup time may be used as an indicator of relative reaction rates. Hence, the greater holdup times required in the moving bed case cause a greater increase in investment than in the fluid bed case where holdup times are lower by a factor of four. Both base case designs are less sensitive to increases in regeneration rate than to decreases, and the reductions in investment which result from doubling the base case holdup times are quite small.

The changes in unit cost produced by variation of the regeneration rate are also very small, as shown in Table 4. Again, the dispersed phase design is more sensitive than the fluid bed design.

The relative holdup factors may be adjusted by changing the inlet and outlet sulfur loadings on the solids, thereby changing the solids circulation rate and also the residence time required. However, for a given set of operating conditions, the effect of holdup on investment and unit cost will decrease as holdup is reduced since the cost of certain items (e.g., internals, cyclones, nozzles, etc.) pertaining to the vessel are essentially constant and will not decrease below a certain level.

It can be seen, then, that the effect of regeneration rate on plant costs is negligible at the flows and rates investigated. Increasing the regeneration rate will produce only slight reductions in plant costs; likewise, decreasing the rate will result in small increases to the plant costs. Unless the regeneration rate can be changed by orders of magnitude, improved regeneration kinetics will not significantly lower either plant investment or unit costs.

2. SORPTION/REGENERATION TEMPERATURES

a. ISOTHERMAL OPERATION AT 600°F

The base case for both the dispersed phase and fluid bed designs includes SO_2 sorption at 600°F (nominal temperature) and sorbent regeneration at 1400°F, while for the fixed bed design the regeneration temperature is 1200°F. In this study, the effects of lowering the sorbent regeneration temperature to 600°F are considered. That is, it has been assumed that the sorbent can be regenerated satisfactorily at 600°F.

For the dispersed phase and fluid bed designs, which will be discussed first, this results in an isothermal sorber - regenerator loop. To make the study as general as possible, heat effects in the sorber and regenerator were not considered, since these will vary with the particular sorbent under consideration. It was also assumed, for the sake of simplicity, that the regeneration kinetics used for the 1400°F base case would apply in the present 600°F case.

The effects of isothermal operation on plant investment are summarized in Tables 5 and 6. These show reductions of $1,925,400 (6.7%) and $951,200 (2.6%) for the dispersed phase and fluid bed designs respectively. The bulk of these reductions occurs in the sorption-regeneration unit, with the air preheater costs also being a significant item. It can be seen that the fluid bed design is less sensitive to this change in regeneration temperature than the dispersed phase design. This is a result of the lower solids flow rate in the fluid bed base case. Hence, the changes in reheat duty, air requirements, etc., are smaller.

In the dispersed phase base case, solids from the regenerator enter the sorber at 1400°F. It was assumed that sorption would not occur until the solids were cooled to approximately 600°F, and sufficient height was added to the sorber to allow for heat transfer between the solids and flue gas. For the present case, no height for heat transfer is needed, since the solids enter the sorber at 600°F, and the decrease in investment for the sorbers reflects this.

In the fluid bed sorber, however, the temperature of the feed solids has essentially no influence on the vessel design since thermal equilibrium between solids and gas occurs rapidly, and no additional bed height need be provided for heat transfer. Hence, lowering the feed solids temperature from 1400°F to 600°F does not change the design.

Since the solids do not have to be heated to regeneration temperature, no solids heater is necessary. As a result, the natural gas and air normally needed to supply this heat are

TABLE 5

EFFECT OF LOWERING REGENERATION TEMPERATURE TO 600°F ON PLANT INVESTMENT

DISPERSED PHASE

	CHANGE IN INVESTMENT ($)[1]
SORBERS	- 336,900
SOLIDS HEATERS	- 562,200
AIR COMPRESSORS	- 512,700
REGENERATORS	- 63,000
AIR-SOLIDS SEPARATORS	+ 49,000
INVENTORY	- 15,700
REGENERATION GAS PLANT	+ 10,000
CLAUS PRETREATMENT	- 38,400
AIR PREHEATER	- 360,000
ELECT. PRECIPITATOR	- 95,500
TOTAL	-1,925,400

PLANT INVESTMENT = $28,790,000[2]

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[1] RELATIVE TO BASE CASE INVESTMENT OF $30,720,000

[2] ROUNDED TO NEAREST $10,000

TABLE 6

EFFECT OF LOWERING REGENERATION TEMPERATURE TO 600°F ON PLANT INVESTMENT

FLUID BED

	CHANGE IN INVESTMENT ($)[1]
SOLIDS HEATERS	- 454,500
AIR COMPRESSORS	- 358,400
REGENERATORS	+ 61,300
AIR-SOLIDS SEPARATORS	+ 24,300
INVENTORY	- 10,500
REGENERATION GAS PLANT	+ 10,000
CLAUS PRETREATMENT	- 38,400
AIR PREHEATER	- 185,000
TOTAL	- 951,200

PLANT INVESTMENT = $34,950,000[2]

[1] RELATIVE TO BASE CASE INVESTMENT OF $35,900,000

[2] ROUNDED TO NEAREST $10,000

eliminated, and the air compressor power requirements are reduced. The elimination of the solids heater and the reduction in size of the air compressor produces the most significant savings in both designs, amounting to $1,074,900 and $812,900 for the dispersed phase and fluid bed cases respectively.

Decreasing the regeneration temperature to 600°F has little effect on investment for the regenerator in either design. Although the smaller volumetric flow of reducing gas, due to the lower temperature, causes a decrease in vessel diameter, the vessel height increases in order to maintain the same solids residence time used in the base case.

In place of the solids heater, an air-solids separator is required. This vessel is necessary in order to prevent the air, which is used to transport the solids from the sorber to the regenerator, from contacting the reducing gas in the regenerator. The required investment for this is very small however.

Regeneration at 600°F requires a 600°F reducing gas fed to the regenerator. This necessitated redesign of the regeneration gas plant and resulted in an increase of $10,000 in investment. The redesigned flowsheet has a different heat exchange scheme wherein high pressure steam is generated in order to produce a reducing gas effluent at 600°F.

There is a small reduction of $38,400 indicated for the Claus Pretreatment. Pretreatment consists of cooling the regenerator effluent to 400°F before feeding it to the Claus plant, and since this effluent is now at 600°F, instead of 1400°F in the base case designs, there is a savings in exchanger costs for this section.

The hot combustion gas effluent from the solids heater normally is added to the main flue gas stream upstream of the preheater, but deletion of the solids heater also eliminates this combustion gas effluent. Consequently the temperature and total flow to the air preheater are reduced, thus causing a significant decrease in investment for the air preheater in both designs. Again, the effect is greater in the dispersed phase design.

This lower total gas flow produces a relatively small decrease in investment for the electrostatic precipitator in the dispersed phase case. In the fluid bed case, the precipitator is unaffected by any process changes since it is upstream of the sorbers.

The effects of "isothermal" operation on operating costs are shown in Tables 7 and 8 for the dispersed phase and fluid bed designs, and indicate, respectively, reductions in unit cost of only $0.37/ton coal (8.2%) and $0.20/ton coal (4.0%). The most important items affected are natural gas costs, heat credits, and capital charges.

Reductions in natural gas charges and power costs are a direct result of the elimination of the solids preheater. Similarly, the heat credit normally taken for the solids heater combustion gas effluent is eliminated. Redesign of the regeneration gas flowsheet produced increased steam credits and a slight reduction in cooling water costs. All other items shown in the table are reduced because of the decrease in plant investment. The overall reduction in unit cost is relatively small for both designs however, with the dispersed phase costs again being more sensitive.

Isothermal operation (@ 600°F) of the fixed bed process produces large savings in plant investment, amounting to nearly $10 million which is equivalent to a 24% reduction. The total change in investment broken down into the various sections is summarized in Table 9 and it is quite obvious that most (~95%) of the reduction occurs in the sorption/regeneration section.

TABLE 7

EFFECT OF LOWERING REGENERATION TEMPERATURE TO 600°F ON OPERATING COST

DISPERSED PHASE

	CHANGE IN OPERATING COST (\$/YR)[1]
MAINTENANCE	- 96,500
SUPPLIES	- 14,500
COOLING WATER	- 4,100
NATURAL GAS	- 250,700
POWER	- 89,900
HEAT CREDIT	-(196,600)[2]
STEAM CREDIT	+ (69,400)[2]
PLANT OVERHEAD	- 55,500
CAPITAL CHARGES	- 250,900
TOTAL	- 634,900

$$\text{CHANGE IN UNIT COST} = \frac{-\$634{,}900/\text{YR}}{1{,}736{,}537\ \text{TONS COAL/YR}} = -\$0.37/\text{TON COAL}$$

---

[1] RELATIVE TO BASE CASE COSTS OF $7,863,200/YR AND $4.53/TON COAL

[2] (   ) INDICATES CREDIT

TABLE 8

EFFECT OF LOWERING REGENERATION TEMPERATURE TO 600°F ON OPERATING COST

FLUID BED

	CHANGE IN OPERATING COST ($/YR)[1]
MAINTENANCE	- 47,500
SUPPLIES	- 7,200
COOLING WATER	- 4,100
NATURAL GAS	- 138,300
POWER	- 50,500
HEAT CREDIT	-(117,800)[2]
STEAM CREDIT	+ (69,400)[2]
PLANT OVERHEAD	- 27,300
CAPITAL CHARGES	- 123,500
TOTAL	- 350,000

$$\text{CHANGE IN UNIT COST} = \frac{-\$350{,}000/\text{YR}}{1{,}736{,}537 \text{ TONS COAL/YR}} = -\$0.20/\text{TON COAL}$$

---

[1] RELATIVE TO BASE CASE COSTS OF $8,646,700/YR AND $4.98/TON COAL

[2] ( ) INDICATES CREDIT

TABLE 9

EFFECT OF LOWERING REGENERATION TEMPERATURE TO 600°F ON PLANT INVESTMENT

FIXED BED

	CHANGE IN INVESTMENT ($) [1]
SORPTION/REGENERATION	-9,190,000
INVENTORY	- 130,000
REGENERATION GAS PLANT	+ 10,000
CLAUS PRETREATMENT	- 90,000
AIR PREHEATER	- 260,000
TOTAL	-9,660,000

PLANT INVESTMENT = $30,550,000

---

[1] RELATIVE TO BASE CASE INVESTMENT OF $40,210,000 [2]

[2] ROUNDED TO NEAREST $10,000

TABLE 10

EFFECT OF LOWERING REGENERATION TEMPERATURE TO 600°F ON OPERATING COST

FIXED BED

	CHANGE IN OPERATING COST ($/YR)[1]
MAINTENANCE	- 483,000
SUPPLIES	- 72,500
COOLING WATER	- 6,700
NATURAL GAS	- 140,200
POWER	- 35,900
HEAT CREDIT	- (169,500)[2]
STEAM CREDIT	(83,600)[2]
PLANT OVERHEAD	- 277,700
CAPITAL CHARGES	- 1,255,800
TOTAL	- 2,185,900

$$\text{CHANGE IN UNIT COST} = \frac{-\$2{,}185{,}900}{1{,}736{,}537 \text{ TONS COAL/YR}} = -\$1.26\text{/TON COAL}$$

---

[1] RELATIVE TO BASE CASE COSTS OF $9,579,500/YR AND $5.52/TON COAL

[2] ( ) INDICATES CREDIT

b. 300°F SORPTION/1400°F REGENERATION

This case considers the effects of lowering the sorption temperature from 600°F, as in the base case, to 300°F. This reduction in sorption temperature is accomplished by relocating the SO_2-removal plant downstream of the power plant air preheater rather than upstream, as it is in the base case. The standard regeneration temperature of 1400°F has been maintained.

The effects of this change on plant investment, for the dispersed phase and fluid bed designs are shown in Table 11. These indicate significantly large reductions of $7,590,000 (24.7%) and $10,210,000 (28.4%) for the two designs respectively.

The bulk of the decrease in investment occurs in the sorption-regeneration unit itself. This is largely owing to the smaller flue gas volume handled at the lower temperature. In the dispersed phase design, there is a substantial decrease in the diameter of the sorbers and in the amount of flue gas ductwork required. In the fluid bed design, there is a similar reduction in flue gas ductowrk, and the number of sorbers is reduced from 10 to 8. Other changes in process equipment are small.

Also shown in the table are cost reductions for the electrostatic precipitator and the power plant air preheater. These cost savings result directly from the relocation of the SO_2-removal plant. In typical power plant operation, flue gas from the power plant is send first to the air preheater and then to the electrostatic precipitator before being vented to the stack. The dispersed phase base case design has the SO_2-removal plant inserted between the power plant and the air preheater, and this results in an incremental investment for both the air preheater and the electrostatic precipitator. Similarly for the fluid bed base case design, there is an incremental investment caused by rearrangement of these units. In this case, the SO_2-removal

TABLE 11

EFFECT OF LOWERING SORPTION TEMPERATURE TO 300°F ON PLANT INVESTMENT

	CHANGE IN INVESTMENT ($)[1] [2]	
	DISPERSED PHASE	FLUID BED
SORPTION/REGENERATION UNIT	- 4,320,000	- 7,140,000
ELECTROSTATIC PRECIPITATOR	- 2,870,000	- 2,870,000
AIR PREHEATER	- 400,000	- 200,000
TOTAL	- 7,590,000	-10,210,000

[1] RELATIVE TO BASE CASE INVESTMENT OF $30,720,000 FOR THE DISPERSED PHASE AND $35,900,000 FOR THE FLUID BED.

[2] ROUNDED TO NEAREST $10,000.

plant is again inserted before the air preheater, but, in addition, the electrostatic precipitator is relocated upstream of the SO_2-removal plant. For the case under consideration here, i.e., 300°F sorption, the typical power plant sequence is undisturbed, since the SO_2-removal plant is "tacked on". Therefore, the base case incremental investments for the air preheater and electrostatic precipitator are eliminated.

The effects of lowering the sorption temperature to 300°F on operating costs are shown in Tables 12 and 13. Reductions in unit cost of $0.81/ton coal (17.9%) for the dispersed phase and $1.23/ton coal (24.7%)for the fluid bed are obtained. It is apparent that the cost reductions are largely the result of the reduced investment, since the three costs which are affected most, viz. maintenance, plant overhead, and capital charges, are a direct function of plant investment.

A significant decrease in heat credits results from the relocation of the SO_2-removal plant. The solids heater effluent which, in the base case, is combined with the sorber effluent and sent to the air preheater, is now vented to stack directly. Consequently, no heat credit can be taken for this.

The other cost changes shown in Tables 12 and 13 are relatively minor. Natural gas costs increase slightly because the lower sorption temperature of 300°F requires a larger reheat duty in the solids heater. Although changes in power costs are negligible for the dispersed phase, significant savings are achieved in the fluid bed design. This results from a substantially larger reduction in horsepower requirements for the flue gas booster fan in the fluid bed case.

For both the dispersed phase and fluid bed designs, the overall savings in plant investment and operating cost achieved by lowering the sorption temperature to 300°F are substantial.

TABLE 12

EFFECT OF LOWERING SORPTION TEMPERATURE TO 300°F ON OPERATING COST

DISPERSED PHASE

	CHANGE IN OPERATING COST ($/YR)[1]
MAINTENANCE	- 379,500
SUPPLIES	- 56,900
NATURAL GAS	+ 57,600
POWER	+ 9,400
HEAT CREDIT	- (176,000)[2]
OVERHEAD	- 218,200
CAPITAL CHARGES	- 986,700
TOTAL	-1,398,300

$$\text{CHANGE IN UNIT COST} = \frac{-\$1,398,300/\text{YR}}{1,736,537 \text{ TONS COAL/YR}} = \$0.81/\text{TON COAL}$$

---

[1] RELATIVE TO BASE CASE COSTS OF $7,863,200/YR AND $4.53/TON COAL

[2] ( ) INDICATES CREDIT

TABLE 13

EFFECT OF LOWERING SORPTION TEMPERATURE TO 300°F ON OPERATING COST

FLUID BED

	CHANGE IN OPERATING COST ($/YR)[1]
MAINTENANCE	- 510,500
SUPPLIES	- 76,600
NATURAL GAS	+ 88,300
POWER	- 135,600
HEAT CREDIT	- (121,700)[2]
PLANT OVERHEAD	- 293,500
CAPITAL CHARGES	-1,327,300
TOTAL	-2,133,500

$$\text{CHANGE IN UNIT COST} = \frac{-\$2,133,500/\text{YR}}{1,736,537\text{TONS COAL/YR}} = -\$1.23/\text{TON COAL}$$

---

[1] RELATIVE TO BASE CASE COSTS OF $8,646,700/YR AND $4.98/TON COAL

[2] ( ) INDICATES CREDIT

This variable is one of the few which has a pronounced effect on plant costs, and illustrates the fact that plant economics are sensitive to changes in volume of flue gas handled by the system. The major factor is the reduced plant investment.

c. 300°F SORPTION/600°F REGENERATION

The effects of lowering both the sorption and regeneration temperatures from the above case levels have been determined for the dispersed phase and fluid bed designs. As in the previously discussed case (VI-B-2-b), the sorption temperature has been lowered to 300°F. However, regeneration is now assumed to occur at 600°F. This case can be viewed as a derivative of the previous case presented in that it enjoys all the economic benefit derived from the use of a low (300°F) sorption temperature. In addition, further cost savings are realized by the use of a low (600°F) regeneration temperature. It should be mentioned, however, that the solids residence time in the regenerator has been assumed to be the same as that used for the 1400°F base case design. Any decrease in regeneration temperature would, of course, increase the solids residence time and thereby minimize the cost savings produced by low temperature regeneration.

Low temperature sorption, as discussed in section VI-B-2-b, substantially reduces the amount of flue gas ducting and the size of the sorber section of the SO_2-removal plant. It also eliminates the incremental costs of the air preheater and the electrostatic precipitator included in the base case designs.

Combining 600°F regeneration with 300°F sorption produces some additional process changes. The size of the solids heater and air compressor are reduced, since the solids reheat duty is much smaller than the base case. The regenerator is reduced in diameter, since it now handles a smaller volume of gas (lower temperature), and there is a corresponding increase in vessel height, based on the assumed solids residence time mentioned above. In order to supply the regenerator with a reducing gas at 600°F, the regeneration gas plant has been redesigned. Finally, the Claus pretreatment exchanger, which is used to cool the regenerator effluent, is reduced in size because of the lower gas temperature.

The effects on plant investment are shown in Table 14 which indicates substantial investment savings of \$8,450,000 (27.5%) and \$10,770,000 (30.0%) for the dispersed phase and fluid bed respectively. It is apparent that the bulk of these cost savings occurs in the sorption-regeneration unit, with the electrostatic precipitator and, to a lesser extent, the air preheater also being affected significantly. Comparison of these figures with those in section IV-B-2-b shows that the major cost changes result from the lower sorption temperature. The added effect of lower regeneration temperature is small.

The effects on plant operating costs are shown in Tables 15 and 16. The reductions in unit cost are substantial, being \$1.13/ton coal (24.9%) and \$1.48/ton coal (29.75) for the dispersed phase and fluid bed designs respectively.

Generally, the savings are similar to those reported in the previous section on low temperature sorption. The most significant additional savings produced by combination of low temperature regeneration with low temperature sorption are in natural gas costs and steam credits. Natural gas costs decrease

TABLE 14

EFFECT OF LOWERING SORPTION/REGENERATION TEMPERATURES TO 300°F/600°F ON PLANT INVESTMENT

	CHANGE IN INVESTMENT ($)[1] [2]	
	DISPERSED PHASE	FLUID BED
SORPTION/REGENERATION UNIT	-5,150,000	- 7,670,000
REGENERATION GAS PLANT	+ 10,000	+ 10,000
CLAUS PRETREATMENT	- 40,000	- 40,000
ELECTROSTATIC PRECIPITATOR	-2,870,000	- 2,870,000
AIR PREHEATER	- 400,000	- 200,000
TOTAL	-8,450,000	-10,770,000

[1] RELATIVE TO BASE CASE INVESTMENT OF $30,720,000 FOR THE DISPERSED PHASE AND $35,900,000 FOR THE FLUID BED.

[2] ROUNDED TO NEAREST $10,000

TABLE 15

EFFECT OF LOWERING SORPTION/REGENERATION TEMPERATURES TO 300°F/600°F ON OPERATING COST

DISPERSED PHASE

	CHANGE IN OPERATING COST (\$/YR)[1]
MAINTENANCE	- 422,500
SUPPLIES	- 63,400
COOLING WATER	- 4,100
NATURAL GAS	- 192,600
POWER	- 80,900
HEAT CREDIT	- (213,700)[2]
STEAM CREDIT	+ (69,400)[2]
PLANT OVERHEAD	- 242,900
CAPITAL CHARGES	-1,098,500
TOTAL	-1,960,600

$$\text{CHANGE IN UNIT COST} = \frac{-\$1{,}960{,}600/\text{YR}}{1{,}736{,}537\ \text{TONS COAL/YR}} = -\$1.13/\text{TON COAL}$$

---

[1] RELATIVE TO BASE CASE COSTS OF $7,863,200/YR AND $4.53/TON COAL

[2] ( ) INDICATES CREDIT

TABLE 16

EFFECT OF LOWERING SORPTION/REGENERATION TEMPERATURES TO 300°F/600°F ON OPERATING COST

FLUID BED

	CHANGE IN OPERATING COST ($/YR)[1]
MAINTENANCE	- 538,500
SUPPLIES	- 80,800
COOLING WATER	- 4,100
NATURAL GAS	- 101,300
POWER	- 187,600
HEAT CREDIT	- (121,700)[2]
STEAM CREDIT	+ (69,400)[2]
PLANT OVERHEAD	- 309,600
CAPITAL CHARGES	-1,400,100
TOTAL	-2,569,700

$$\text{CHANGE IN UNIT COST} = \frac{-\$2{,}569{,}700/\text{YR}}{1{,}736{,}537 \text{ TONS COAL/YR}} = -\$1.48/\text{TON COAL}$$

[1] RELATIVE TO BASE CASE COSTS OF $8,646,700/YR AND $4.98/TON COAL

[2] ( ) INDICATES CREDIT

because of the lower solids reheat duty. Additional steam credits resulted from redesign of the regeneration gas plant. Overall, the most important cause of the reduction in unit cost is the large decrease in investment, for both process designs.

3. SO_2-REMOVAL EFFICIENCY

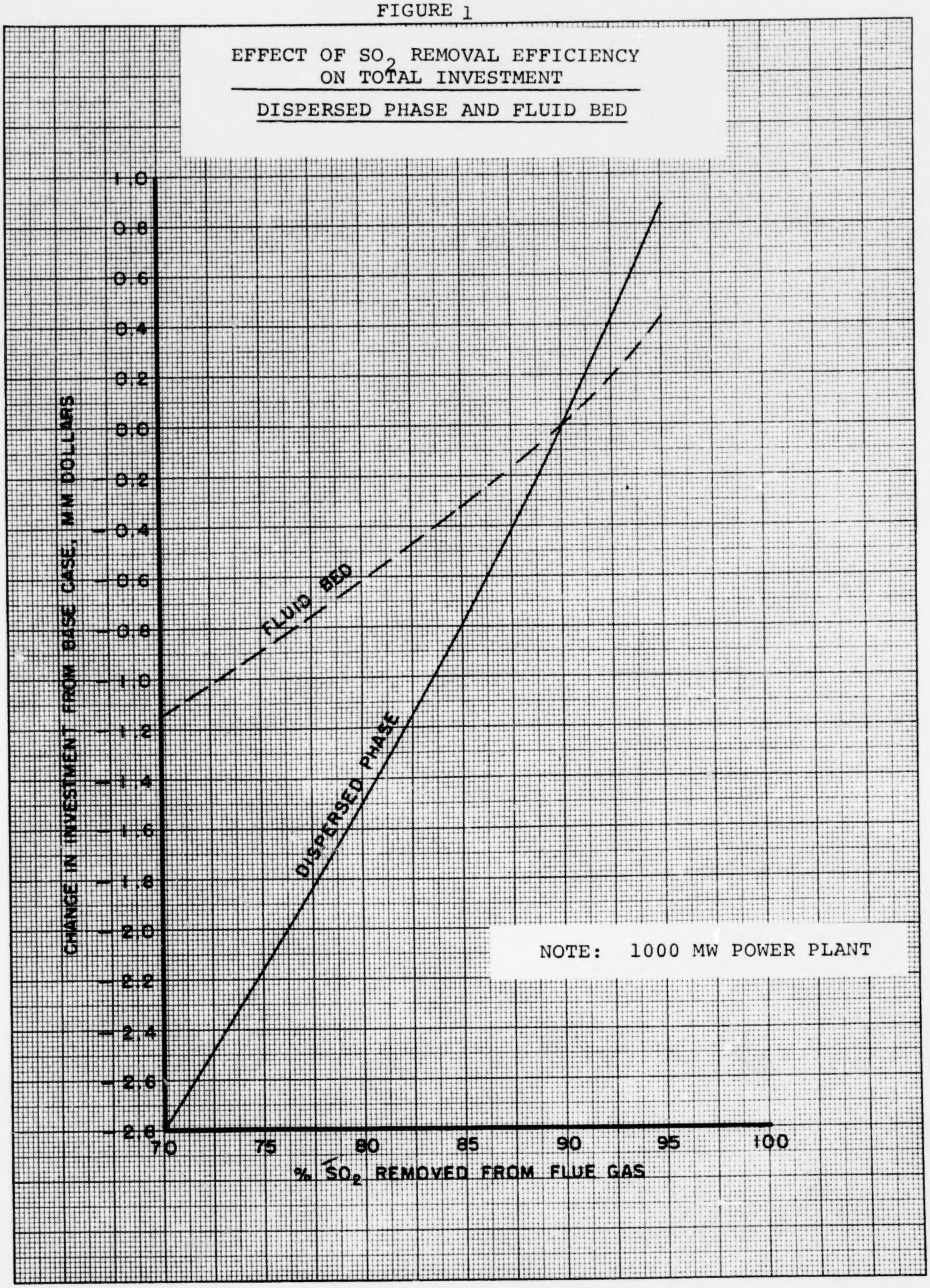
The base case for both the dispersed phase and fluid bed designs assumes 90% removal of the SO_2 in the power plant flue gas. This study is concerned with the effects on plant economics of varying the amount of SO_2 removed from the flue gas. SO_2-removal efficiencies of 70% and 95% have been compared to the base case.

The effects of SO_2-removal efficiency on investment are shown in Figure 1 for the dispersed phase and fluid bed designs respectively. Comparison of these curves shows that the dispersed phase design is much more sensitive to changes in the amount of SO_2 removed than the fluid bed. For SO_2-removal efficiencies of 70% and 95%, the changes in investment are -\$2,800,000 (-9.1%) and +\$880,000 (+2.9%) for the dispersed phase, and -\$1,150,000 (-3.2%) and +\$430,000 (+1.2%) for the fluid bed.

A reduction of SO_2-removal from 90% to 70% decreases the solids circulation rate and reheat duty and lowers the amount of sulfur handled by the system. The effects are more pronounced in the dispersed phase design since the initial (base case) solids circulation rate is greater than that in the fluid bed design.

In both designs, substantial savings in investment are realized by reductions in size of the air compressor, regeneration gas plant, and the Claus plant. These savings total \$1,080,000 and \$980,000 for the dispersed phase and fluid bed respectively. The air compressor decreases in cost since the smaller flow of solids requires less air for both transport and combustion in the solids heater. The reduction in investment for the regeneration gas plant and the Claus plant are a direct result of the smaller amount of sulfur being handled. Less reducing gas is required and less sulfur is produced.

FIGURE 1



EFFECT OF SO_2 REMOVAL EFFICIENCY ON TOTAL INVESTMENT

DISPERSED PHASE AND FLUID BED

NOTE: 1000 MW POWER PLANT

There are three other items which are affected significantly only in the dispersed phase design. First, there is a decrease of $300,000 in investment for the dispersed phase sorbers, since less SO_2 is removed from the flue gas. In the fluid bed case, the minimum sorber bed height was established not by kinetics, which requires a small bed height, but by hydrodynamic considerations and operability as a fluid bed unit. Hence, no reduction in bed height is possible. Second, a large investment savings of $1,000,000 for the electrostatic precipitator results from the reduction in SO_2-removal efficiency. The efficiency of the electrostatic precipitator, which is downstream of the SO_2-removal plant in the dispersed phase design, is influenced by the amount of sulfur present in the gas, being more efficient with greater sulfur concentrations. In the 70% SO_2-removal case, the sulfur content of the gas to the precipitator is three times that in the base case (90% SO_2-removal) and results in a decrease in size of the precipitator through increased efficiency. The fluid bed design is not influenced since the electrostatic precipitator is upstream of the SO_2-removal plant. Third, there is a decrease of about $170,000 in air preheater costs because of the lower total gas flow which results from the decrease in transport and combustion air used in the process. This effect is present in the fluid bed design also, but is not as pronounced because of the smaller changes in process air requirements.

There are other decreases in investment, but their effects are relatively minor. These include the solids heater, inventory, and Claus pretreatment.

An increase in SO_2 removal from 90% to 95% produces the opposite effects to those cited above, but changes are smaller. The most significant increases in investment occur in the regeneration gas plant, for both designs, and in the electrostatic precipitator for the dispersed phase.

For the fluid bed sorber, the bed height calculated from kinetics is, again, less than the minimum bed height established for the base case. Hence, no change in sorber investment occurs. It should be noted, however, that above the point at which this minimum bed height is predicted by kinetics, investment and operating costs will increase rapidly with increasing SO_2-removal efficiency since the pressure drop through the sorber bed will increase with a resulting, and substantial, increase in the size of the flue gas booster fans.

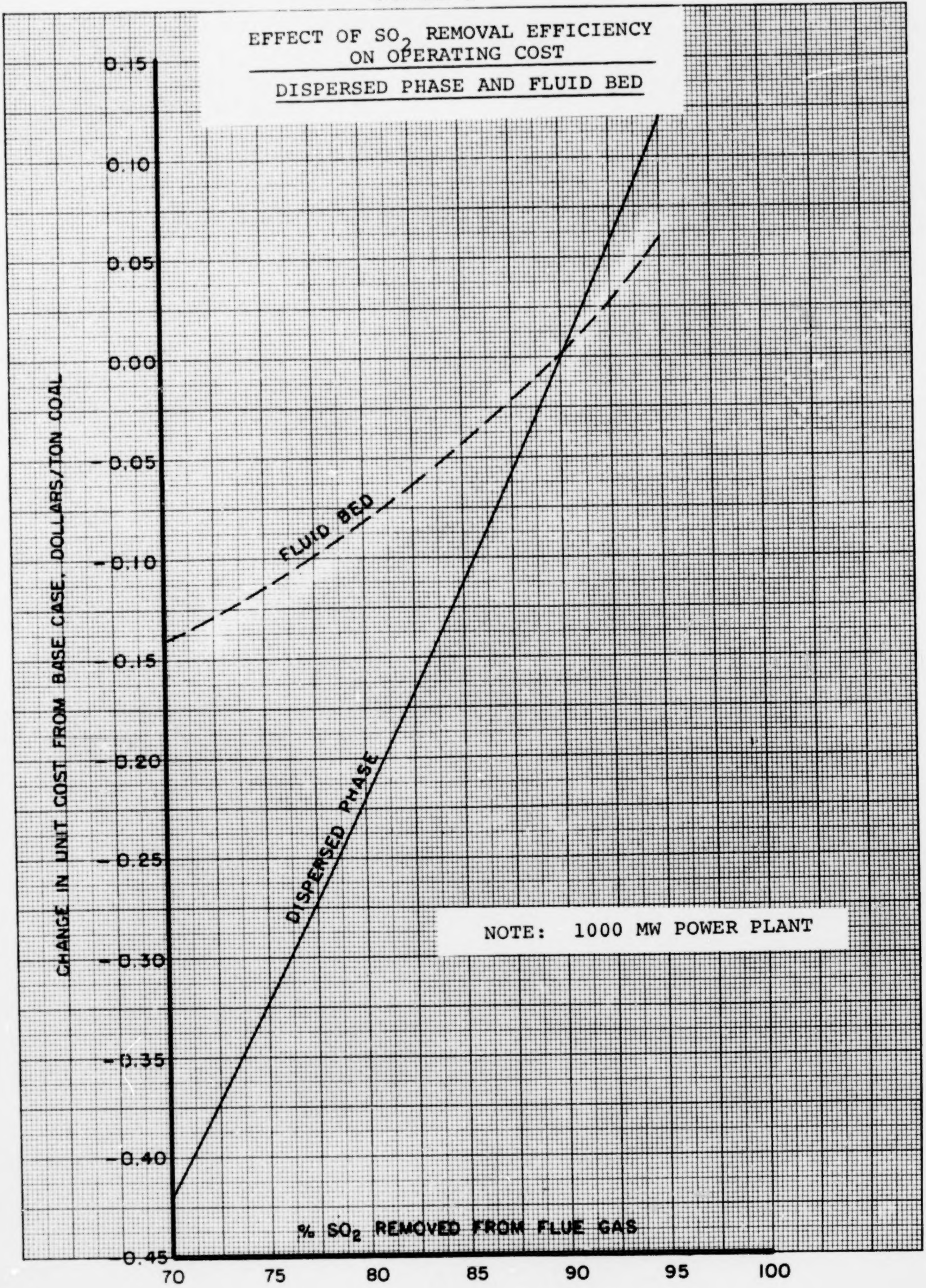
The effects of SO_2-removal efficiency on unit cost are shown in Figure 2. These show, for 70% and 95% SO_2-removal efficiencies, changes in unit cost of -$0.42/ton coal and +$0.12/ton coal for the dispersed phase and -$0.14/ton coal and +$0.06/ton coal for the fluid bed.

The most significant changes in operating costs are in natural gas charges, sulfur credits, and capital charges. Natural gas charges change as a result of the changes in solids reheat duty. Varying the amount of SO_2 recovered from the flue gas produces differences in the sulfur credits. The changes in capital charges, of course, are due to the changes in plant investment.

In addition to these, the change in attrition losses is an important item in the dispersed phase design. This is due to the high base case solids circulation. For example, reduction of the amount of SO_2 removed from 90% to 70% decreases the attrition losses by about $200,000/yr or $0.12/ton coal.

Again, the dispersed phase is the more sensitive of the two process designs. However, neither design shows very much change in unit cost over the range of SO_2 removal efficiencies investigated, the maximum being less than 10% for the dispersed phase design at 70% SO_2-removal.

FIGURE 2


EFFECT OF SO_2 REMOVAL EFFICIENCY ON OPERATING COST
DISPERSED PHASE AND FLUID BED
NOTE: 1000 MW POWER PLANT
CHANGE IN UNIT COST FROM BASE CASE, DOLLARS/TON COAL
% SO_2 REMOVED FROM FLUE GAS
FLUID BED
DISPERSED PHASE
0.15
0.10
0.05
0.00
-0.05
-0.10
-0.15
-0.20
-0.25
-0.30
-0.35
-0.40
-0.45
70
75
80
85
90
95
100

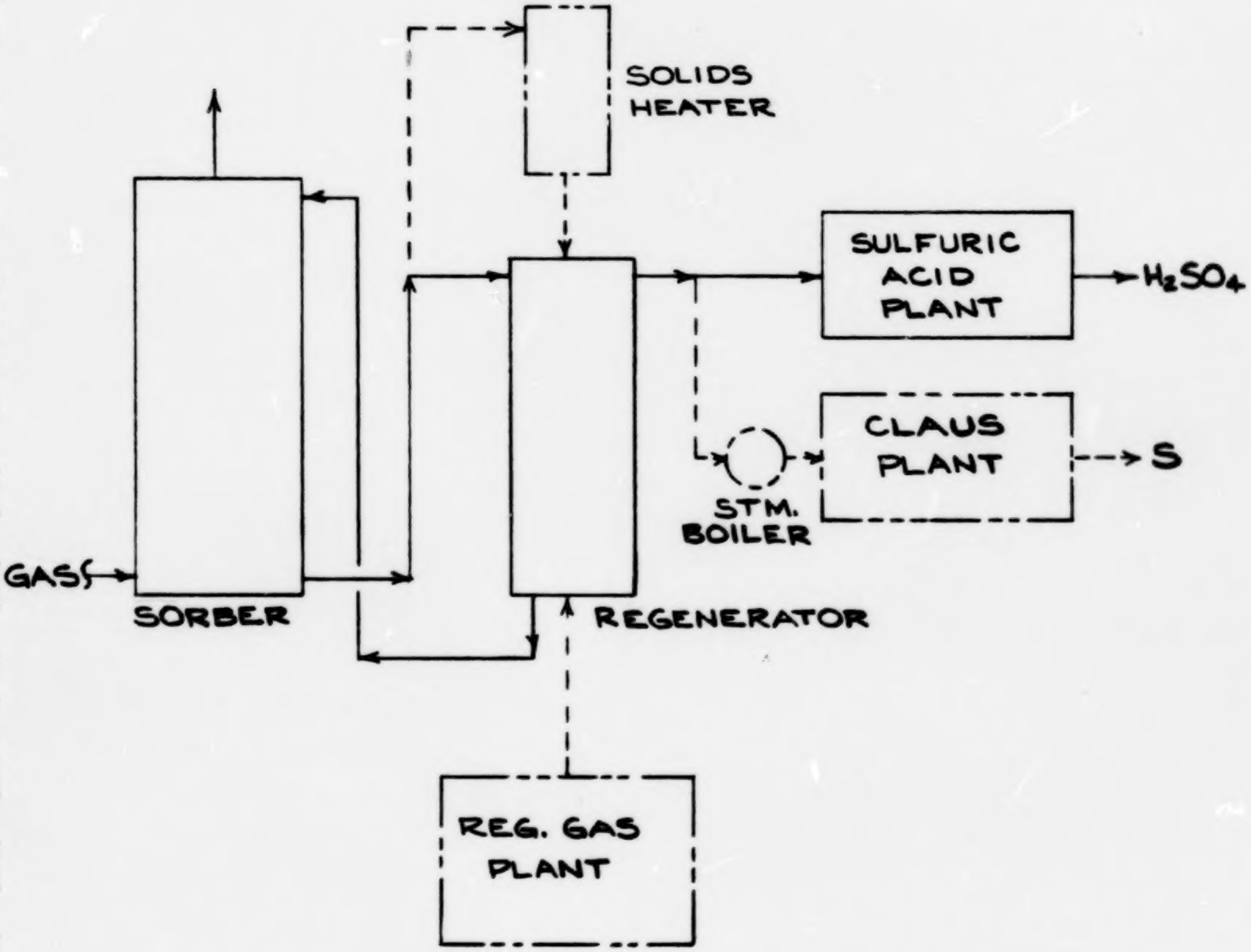
4. THERMAL REGENERATION

In the base case designs, the sorbent is chemically regenerated at 1400°F using reducing gas to recover sulfur in the form of H_2S which is then sent to a Claus unit where the H_2S is converted to elemental sulfur. This evaluation, for the dispersed phase and fluid bed designs, assumes that the sorbent can be regenerated satisfactorily by heating to 1400°F in the presence of air to form an SO_2-rich effluent. Heat for this thermal regeneration is supplied by direct combustion of natural gas in the sorbent bed. The SO_2-rich regenerator off-gas is sent to a sulfuric acid plant where 98% H_2SO_4 is produced as a salable by-product.

Because of lack of complete data, particularly kinetics, on thermally regenerable sorbents, design parameters were assumed for an unspecified metal oxide sorbent (e.g., copper oxide). Sorbent loadings were maintained at the same levels used in the base case designs to keep the solids circulation rates unchanged. Heat effects and kinetics in the sorbers were also assumed to be the same as in the base case designs, resulting in no change in size of the sorbers. The heat of reaction for the decomposition of copper sulfate was used in regeneration calculations, assuming all desorbed sulfur is removed as SO_2. For both process designs, a 30 minute solids residence time was assumed adequate to regenerate the sorbent.

Thermal regeneration eliminates the need for a regeneration gas plant since no reducing gas is required. The regenerator can also be deleted by increasing the size of the solids heater so that it can be used not only to heat the solids to 1400°F, but also to regenerate them. A sulfuric acid plant is substituted for the Claus plant. A schematic flowsheet which shows this processing scheme is shown in Figure 3.

FIGURE 3


SOLIDS HEATER
SULFURIC ACID PLANT
H_2SO_4
CLAUS PLANT
S
STM. BOILER
GAS
SORBER
REGENERATOR
REG. GAS PLANT

NOTE: DASHED LINES INDICATE EQUIPMENT NOT REQUIRED FOR THERMAL REGENERATION

SCHEMATIC FLOW SHEET FOR THERMAL REGENERATION

In addition to the process changes mentioned above, the size of the air compressor and power plant air preheater are affected, as is the electrostatic precipitator in the dispersed phase design. Since heat must be supplied not only to raise the solids to regeneration temperature but also to regenerate them, there is an increase in combustion air, with a consequent increase in size of the air compressor. The effluent gas from the solids heater is normally combined with the cleaned flue gas from the sorber and sent through the air preheater, but in the case of thermal regeneration the entire effluent from the heater-regenerator is sent to a sulfuric acid plant. Consequently, the total flow of gas through the air preheater is reduced and this decreases the size of the preheater and, in the dispersed phase design, the electrostatic precipitator which is downstream of the sorber.

The effects of these changes on plant investment are shown in Tables 17 and 18 for the dispersed phase and fluid bed designs respectively. Modest savings of $1,830,000 (6.0%) for the dispersed phase and $1,040,000 (2.9%) for the fluid bed are indicated.

The major contributions to the cost changes, for both designs, result from the deletion of the regeneration gas plant and Claus plant and the addition of a sulfuric acid plant. The first two are virtually an economic standoff with the effect that the Claus plant essentially represents the cost savings realized by substitution of thermal for chemical regeneration.

It should be noted that a rather large reduction in inventory is shown for the dispersed phase. This is caused by the reduced solids residence time needed for regeneration (30 minutes vs. 2 hours in the base case). Although this residence time has been assumed, a significant reduction from the base case residence time should be achieved by the use of thermal regeneration in a fluid bed vessel rather than chemical regeneration in a moving bed unit.

TABLE 17

EFFECT OF THERMAL REGENERATION ON PLANT INVESTMENT

DISPERSED PHASE

	CHANGE IN INVESTMENT ($)[1]
SORPTION/REGENERATION UNIT	+ 88,400
INVENTORY	- 374,000
REGENERATION GAS PLANT	-3,300,000
CLAUS PRETREATMENT	- 102,600
CLAUS PLANT	-1,400,000
SULFURIC ACID PLANT	+3,640,000
AIR PREHEATER	- 184,000
ELECTROSTATIC PRECIPITATOR	- 195,000
TOTAL	-1,827,200

PLANT INVESTMENT = $28,890,000[2]

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[1] RELATIVE TO BASE CASE INVESTMENT OF $30,720,000

[2] ROUNDED TO NEAREST $10,000

TABLE 18

EFFECT OF THERMAL REGENERATION ON PLANT INVESTMENT

FLUID BED

	CHANGE IN INVESTMENT ($)[1]
SORPTION/REGENERATION UNIT	+ 199,700
INVENTORY	- 24,900
REGENERATION GAS PLANT	-3,300,000
CLAUS PRETREATMENT	- 102,600
CLAUS PLANT	-1,400,000
SULFURIC ACID PLANT	+3,640,000
AIR PREHEATER	- 50,000
TOTAL	-1,037,800

PLANT INVESTMENT = $34,860,000[2]

[1] RELATIVE TO BASE CASE INVESTMENT OF $35,900,000

[2] ROUNDED TO NEAREST $10,000

Final recovery of sulfur as concentrated H_2SO_4 instead of elemental sulfur would probably require additional expenditures for costlier handling and storage facilities. However, for this study, no such additional costs were included.

Reductions in operating costs which result from the use of thermal regeneration are shown in Tables 19 and 20. These show decreases in unit costs of $0.52/ton coal (11.5%) and $0.44/ton coal (8.8%) for the dispersed phase and fluid bed designs respectively. Both process designs exhibit about the same degree of cost sensitivity to thermal regeneration.

The single most important contribution to the decrease in unit cost is the natural gas cost. Although the natural gas consumption increases in the solids heater-regenerator in order to supply heat for thermal regeneration, there is a large reduction caused by the elimination of the regeneration gas plant.

The increase in sulfur credit is also significant. In this case the sulfur is credited as concentrated (98%) sulfuric acid at $8/ton H_2SO_4 as against elemental sulfur in the base case. This unit credit corresponds to the base case credit of $20/ton sulfur, with an additional processing cost of about $1.50/ton H_2SO_4 added.

Capital charge changes are also important in adding to the decrease in unit cost. Naturally, these result from the reduction in plant investment. The other items listed are relatively small.

Overall, the reductions in costs achieved through the use of thermal regeneration are relatively small for both process designs. In addition, there is a potential disadvantage to this type of operation, viz, the need to dispose of sulfuric acid rather than elemental sulfur, with a strong possibility that the acid will be contaminated with trace components such as NO_x, fly ash, etc.

TABLE 19

EFFECT OF THERMAL REGENERATION ON OPERATING COST

DISPERSED PHASE

	CHANGE IN OPERATING COST ($/YR)[1]
LABOR & SUPERVISION	- 35,200
MAINTENANCE & SUPPLIES	- 105,200
COOLING WATER	+ 41,400
NATURAL GAS	- 549,400
POWER	+ 92,100
HEAT CREDITS	-(109,700)[2]
STEAM CREDITS	- (44,100)[2]
PAYROLL & PLANT OVERHEAD	- 77,200
CAPITAL CHARGES	- 237,900
SULFUR CREDIT[3]	(187,700)[2]
TOTAL	- 905,300

$$\text{CHANGE IN UNIT COST} = \frac{-\$905{,}300/\text{YR}}{1{,}736{,}537\ \text{TONS COAL/YR}} = -\$0.52/\text{TON COAL}$$

[1] RELATIVE TO BASE CASE COSTS OF $7,863,200/YR AND $4.53/TON COAL

[2] ( ) INDICATES CREDIT

[3] DIFFERENCE BETWEEN SULFURIC ACID CREDIT, AT $8/TON H_2SO_4, AND BASE CASE SULFUR CREDIT, AT $20/TON S.

TABLE 20

EFFECT OF THERMAL REGENERATION ON OPERATING COST

FLUID BED

	CHANGE IN OPERATING COST ($/YR)[1]
LABOR & SUPERVISION	- 35,300
MAINTENANCE & SUPPLIES	- 59,800
COOLING WATER	+ 41,400
NATURAL GAS	- 553,500
POWER	+ 90,100
HEAT CREDITS	- (95,500)[2]
STEAM CREDITS	- (44,100)[2]
PAYROLL & PLANT OVERHEAD	- 54,600
CAPITAL CHARGES	- 135,200
SULFUR CREDIT[3]	(187,700)
TOTAL	- 755,000

$$\text{CHANGE IN UNIT COST} = \frac{-\$755{,}000/\text{YR}}{1{,}736{,}537 \text{ TONS COAL/YR}} = -\$0.44/\text{TON COAL}$$

---

[1] RELATIVE TO BASE CASE COSTS OF $8,646,700/YR AND $4.98/TON COAL

[2] (   ) INDICATES CREDIT

[3] DIFFERENCE BETWEEN SULFURIC ACID CREDIT, AT $8/TON H_2SO_4, AND BASE CASE SULFUR CREDIT, AT $20/TON S.

C. INVESTMENT

1. FIXED CAPITAL INVESTMENT, LOAD FACTOR, CAPITAL CHARGE RATE

About 50% of the total annual costs computed for the three processes studied is directly a function of capital investment. For a given capital investment and plant capacity these "capital charges" will vary with load factor[1] and capital charge rate. Because of the importance of capital charges to process economics, relationships permitting rapid evaluation of cost changes relative to the base case are presented in Figures 4 through 10. Thus the change in cost, expressed as dollars per ton of power plant coal burned, resulting from changes in investment can be obtained from Figures 4-10 and added to the base case cost for each process to obtain alkalized alumina plant costs for any combination of load factor and capital charges falling within the range studied.

The change in operating cost as a function of load factor is shown in Figure 4 for each of the three base cases; i.e., dispersed phase, fluid bed, and fixed bed. As is evident from Figure 4 the dispersed phase operating cost is affected the least by load factor while the fixed bed shows the greatest effect. This is not unexpected since the fixed bed has the highest investment and capital charges are sensitive to load factor whereas other costs (e.g., utilities) are not.

Figures 4-7 show the unit cost change as a function of load factor and capital investment for a single capital charge rate, viz, 13% of investment per year. The effect of other charge rates on operating costs can be determined from Figures 8 (base cases) and 9, 10. It should be noted that for any given investment the portion

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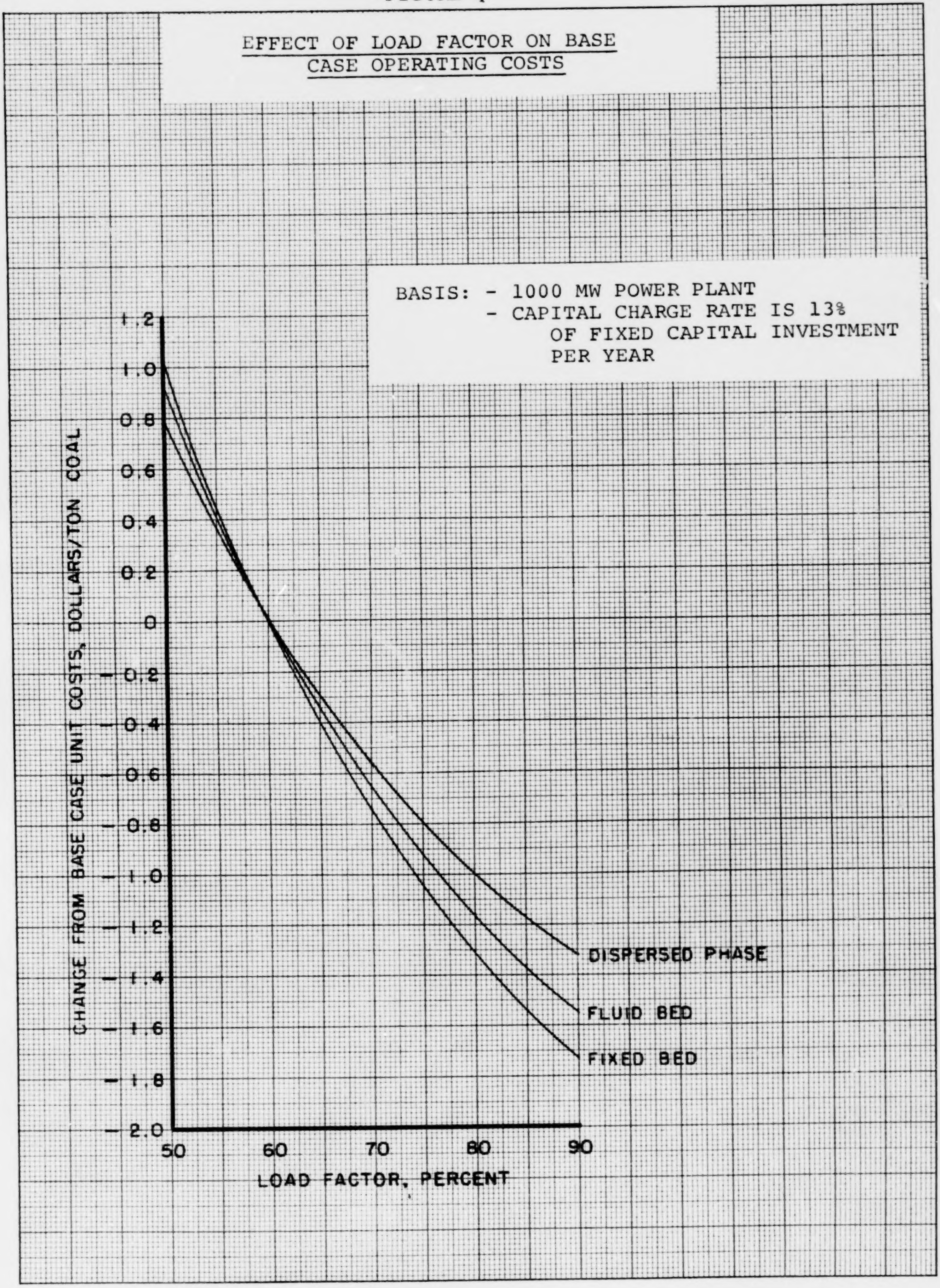
[1] See Section E.3 for discussion of load factor.

FIGURE 4



EFFECT OF LOAD FACTOR ON BASE CASE OPERATING COSTS

BASIS: - 1000 MW POWER PLANT
- CAPITAL CHARGE RATE IS 13% OF FIXED CAPITAL INVESTMENT PER YEAR

FIGURE 5

EFFECT OF LOAD FACTOR ON OPERATING COSTS FOR VARIOUS FIXED CAPITAL INVESTMENTS DISPERSED PHASE

BASIS: - 1000 MW POWER PLANT
- CAPITAL CHARGE RATE TAKEN AS 13% FIXED CAPITAL INVESTMENT/YEAR

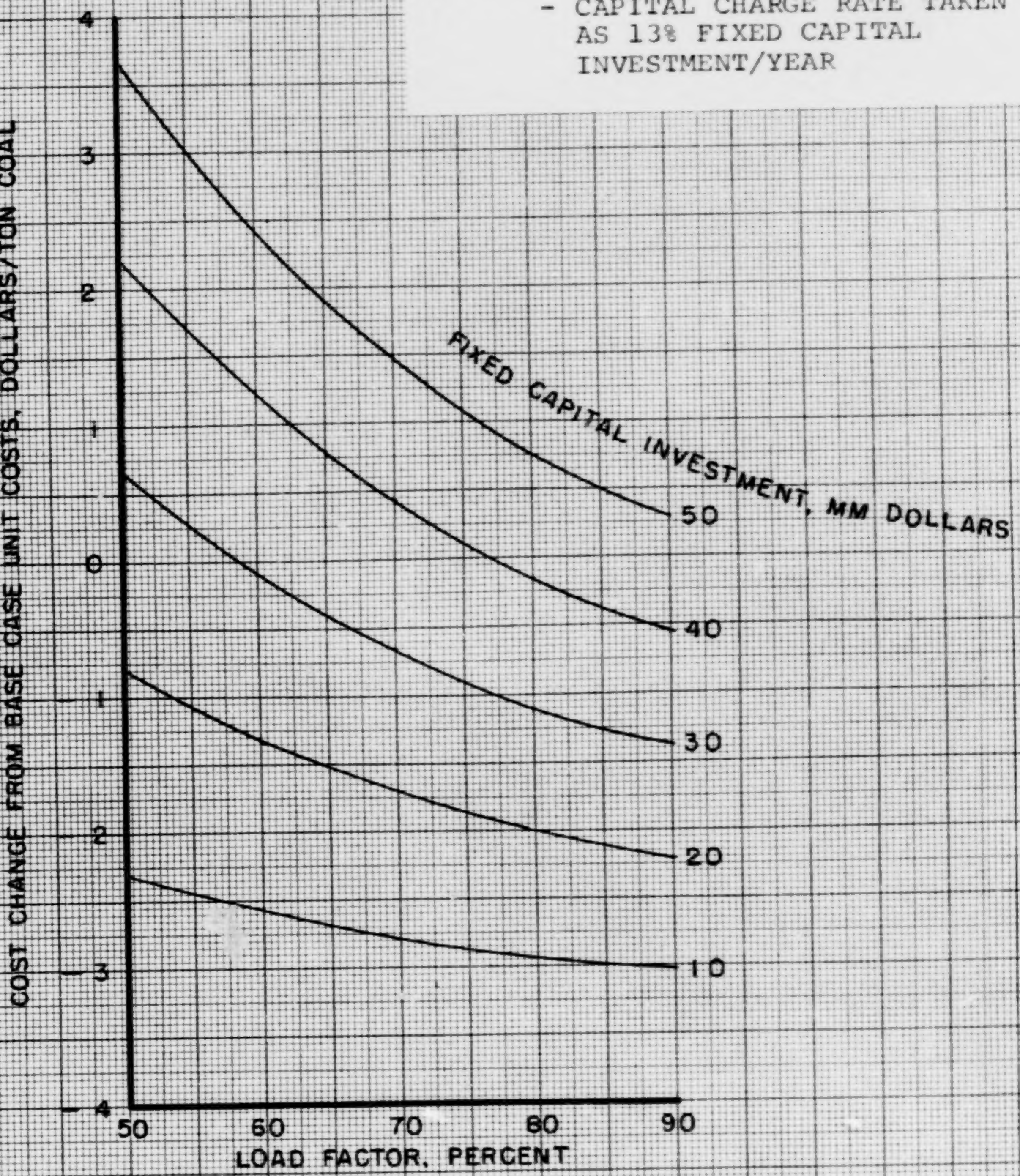
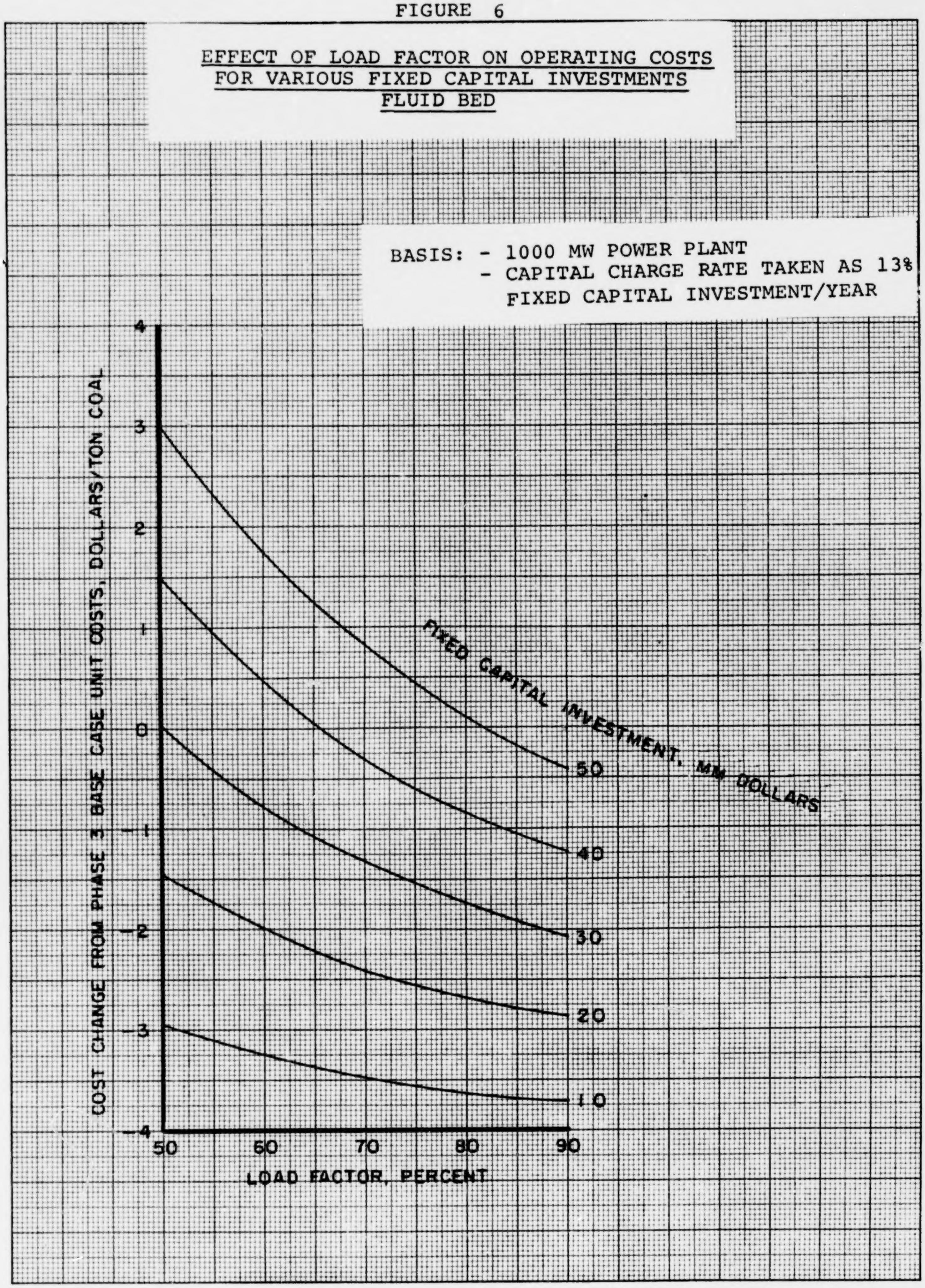


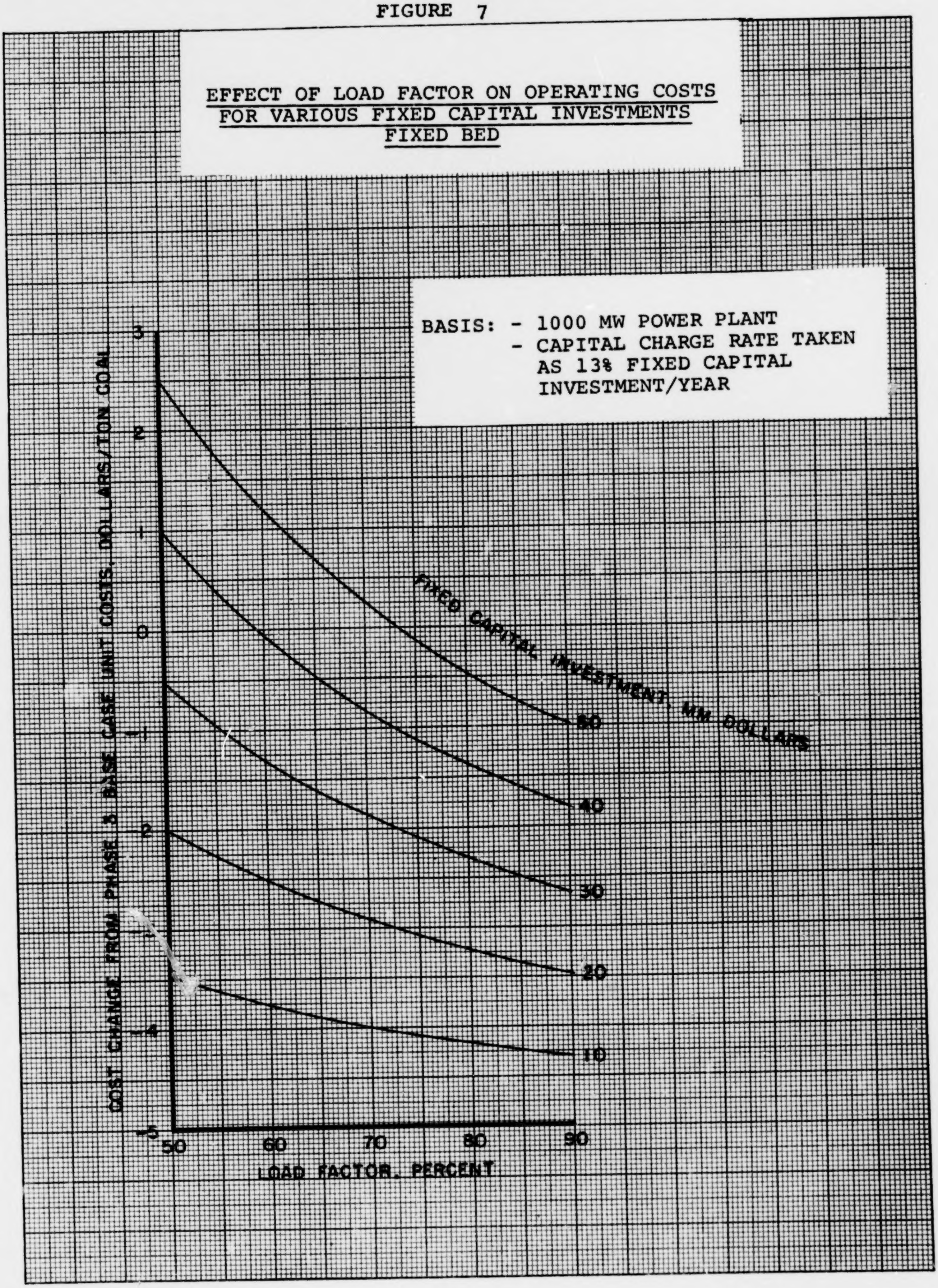
FIGURE 6



EFFECT OF LOAD FACTOR ON OPERATING COSTS
FOR VARIOUS FIXED CAPITAL INVESTMENTS
FLUID BED

BASIS: - 1000 MW POWER PLANT
- CAPITAL CHARGE RATE TAKEN AS 13% FIXED CAPITAL INVESTMENT/YEAR

FIGURE 7



EFFECT OF LOAD FACTOR ON OPERATING COSTS FOR VARIOUS FIXED CAPITAL INVESTMENTS FIXED BED

BASIS: - 1000 MW POWER PLANT
- CAPITAL CHARGE RATE TAKEN AS 13% FIXED CAPITAL INVESTMENT/YEAR

FIGURE 8

EFFECT OF CAPITAL CHARGES ON OPERATING COSTS

BASIS: - 1000 MW POWER PLANT
- 60% LOAD FACTOR
- BASE CASE UNIT COSTS: $/TON

	$/TON
DISPERSED PHASE	4.53
FLUID BED	4.98
FIXED BED	5.52

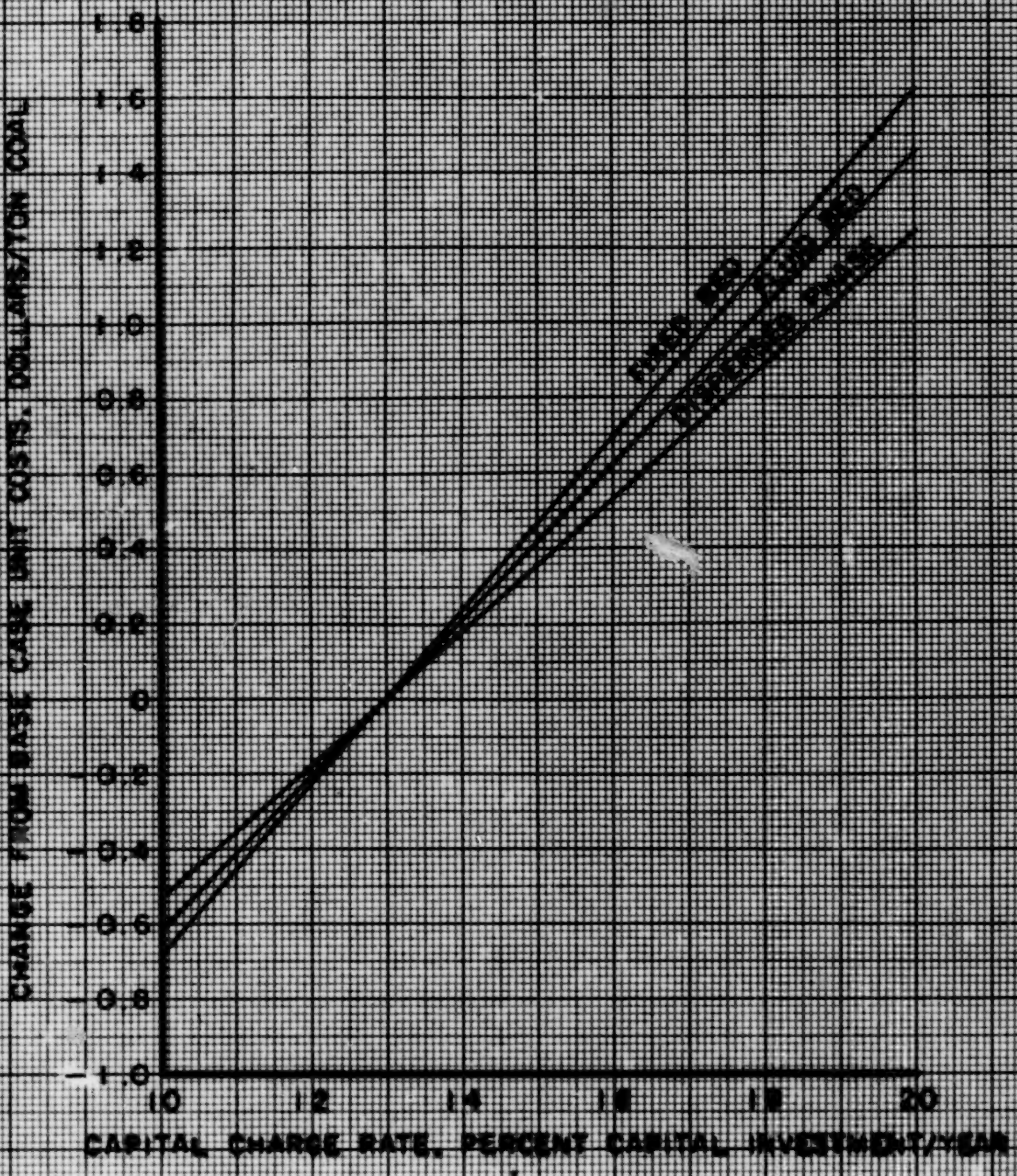
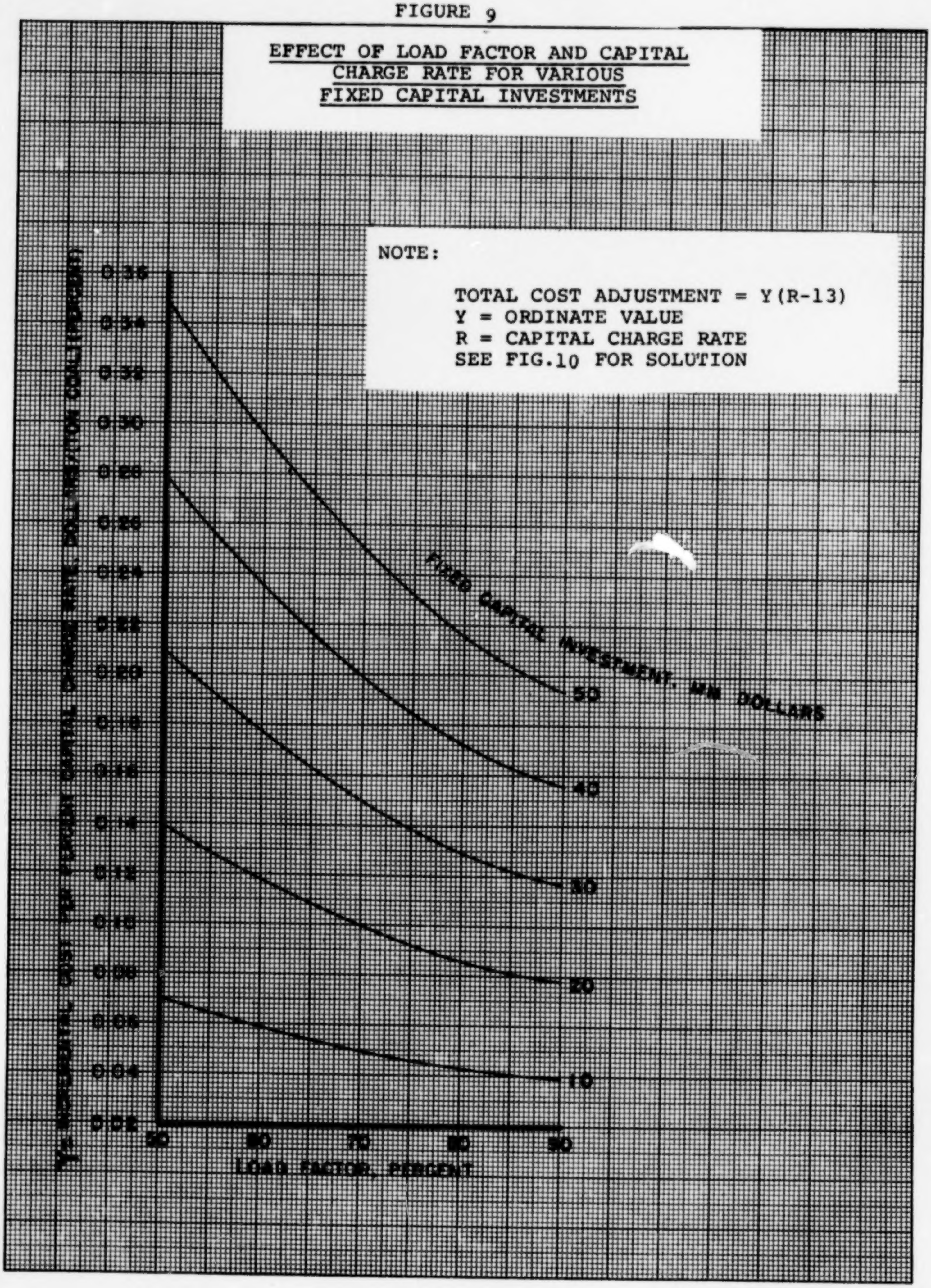


FIGURE 9

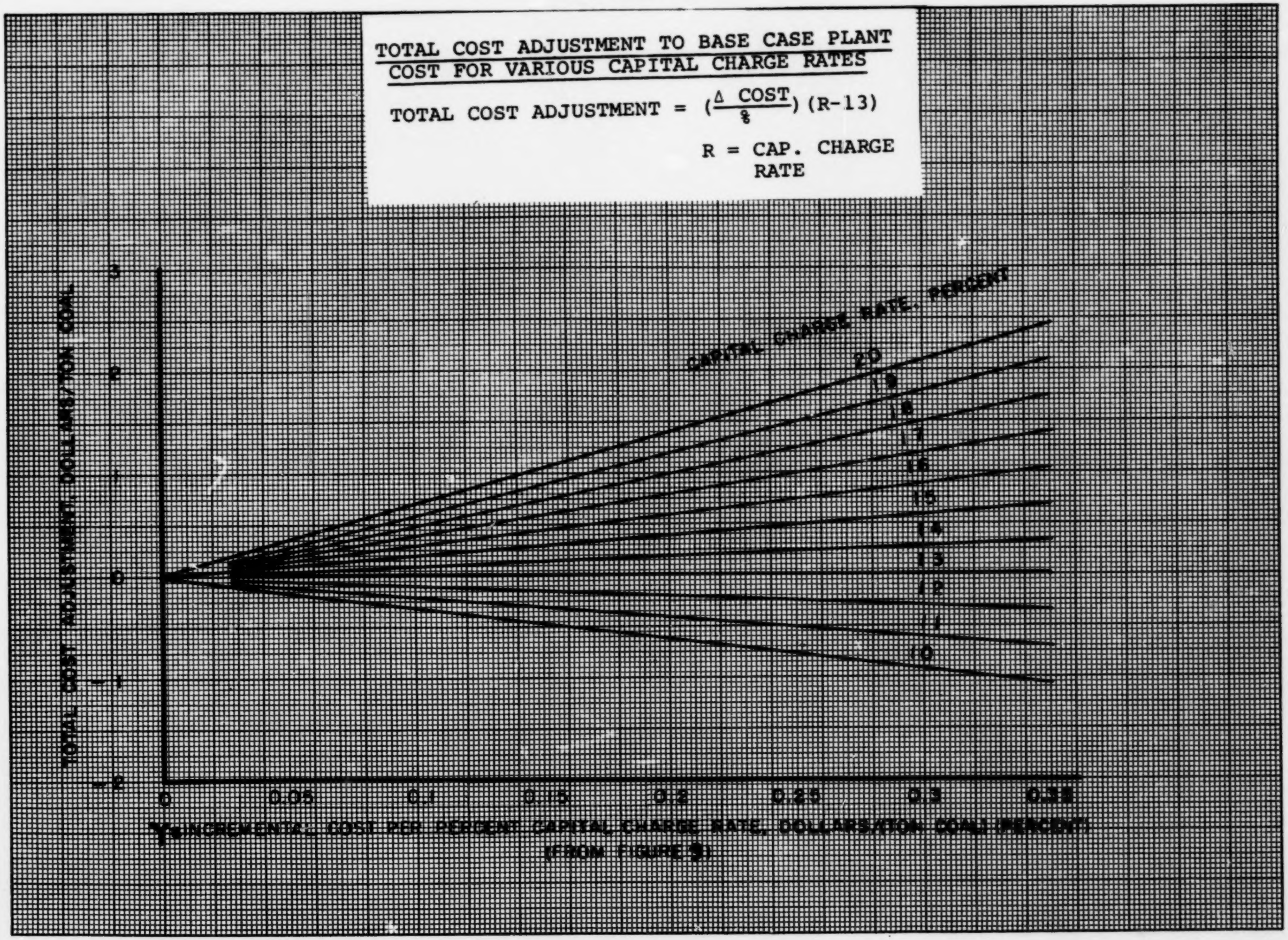
EFFECT OF LOAD FACTOR AND CAPITAL CHARGE RATE FOR VARIOUS FIXED CAPITAL INVESTMENTS



NOTE:

TOTAL COST ADJUSTMENT = Y(R-13)
Y = ORDINATE VALUE
R = CAPITAL CHARGE RATE
SEE FIG.10 FOR SOLUTION

FIGURE 10


TOTAL COST ADJUSTMENT TO BASE CASE PLANT COST FOR VARIOUS CAPITAL CHARGE RATES

TOTAL COST ADJUSTMENT = $\left(\frac{\Delta\ \text{COST}}{\%}\right)(R-13)$

R = CAP. CHARGE RATE

of the total cost owing to capital charge rate is independent of process design and therefore, by calculating incremental rather than total costs, a separate chart for each process is not required.

As is evident from Figures 4-7, the effect of load factor on plant costs increases as the plant investment increases. For example, at a $10,000,000 investment, increasing the load factor from 50 to 90% decreases the plant cost by about $0.75/ton coal compared to a decrease of about $3.45/ton at a $50,000,000 investment. For the same change in investment, i.e., from 10 to 50 million dollars, the plant costs change about $3.30/ton coal at 90% load factor compared to about $6.00/ton of coal at 50%. The cost change is actually a linear function of investment at any given load factor, ranging from about $1.50/ton per $10,000,000 at 50% to about $0.85/ton per $10,000,000 at 90%. At the base case load factor of 60% the cost change is about $1.25/ton coal per $10,000,000 change in investment. Thus if the base case load factor were increased from 60% to 90%, the unit costs would decrease by about $0.4/ton coal per 10 million dollars of plant investment, or for a 30 million dollar plant a total reduction of about $1.20/ton coal would be obtained.

The base case costs in dollars per ton of coal are 4.53, 4.98, and 5.52 for the dispersed phase, fluid bed, and fixed bed designs respectively. (The above values include attrition loss as specified for each case, sulfur credit @ $20/short ton, a natural gas price of 27¢/MM Btu, and a stream efficiency of 60%.) An alkalized alumina operating cost of $2.00/ton coal, which has been mentioned as a reasonable goal, would limit the investment to about $10,000,000 at base case conditions or to about $15,000,000 at 90% load factor. It appears, therefore, that the SO_2-removal plant investment must be reduced by a factor of about 2 to 3 to achieve costs in the order of $2/ton of power plant coal burned.

As previously mentioned, Figures 4-7 are based on a capital charge rate of 13%. The effect of different charge rates can be determined from Figures 8, 9 and 10. Figure 10 shows the incremental cost per percent capital charge rate as a function of load factor and plant investment. Since each of the process designs being studied has a different base case cost, a separate set of curves for each case would be required if actual plant costs were shown. The effect of load factor and capital charge rate on plant costs are dependent only on plant investment, however, and not on process design so a single curve for each investment level is adequate for all cases when an incremental basis is used. Therefore once the incremental cost, which is the same for all three designs at equivalent conditions, is determined it can be added to the appropriate base costs to obtain actual plant costs.

From Figure 9 the incremental cost per percent capital charge rate is obtained for any desired load factor and investment, and the total cost adjustment (TCA) is calculated by the equation TCA = Y(R-13), where Y = Figure 9 ordinate value, and R = capital charge rate.

To facilitate calculations, the solution to the above equation is given in Figure 10. To use, the abscissa of Figure 10 is entered at the ordinate value read from Figure 9 and a vertical line is projected upward to the desired capital charge rate line, thence horizontal to the ordinate where the total cost adjustment to plant cost is read directly. Since the base case costs are based on a 13% charge rate, the adjustment is zero for this rate, negative for rates less than 13%, and positive for rates greater than 13%.

The capital charge rate is a function of plant life, method of depreciation, cost of money, and taxes and insurance. The base case rate of 13% was based on NAPCA's suggested value of a 10 year straight line depreciation (i.e., 10%) with 3% added for taxes and insurance. A more detailed breakdown of capital charges is given by TVA in a paper presented at the Second Semiannual Control Process Contractors' Meeting in Cincinnati, Ohio, June 11-13, 1969. This paper shows annual capital charges - levelized basis of 14.5% based on a 35-year life and straight line depreciation. In this same paper, TVA states that capital charges, when levelized to an average annual basis, usually run between 13 and 16% of the original investment. Thus, it is felt that a method of evaluating capital charge rates between 10 and 20% as is provided with Figures 8, 9, and 10, should adequately cover the range need for SO_2 removal processes. That is, the capital charge rate corresponding to the desired plant life, cost of money, and taxes and insurance is calculated and then used with Figures 8 or 9 and 10 to determine the change in plant costs that should be added to the base case costs.

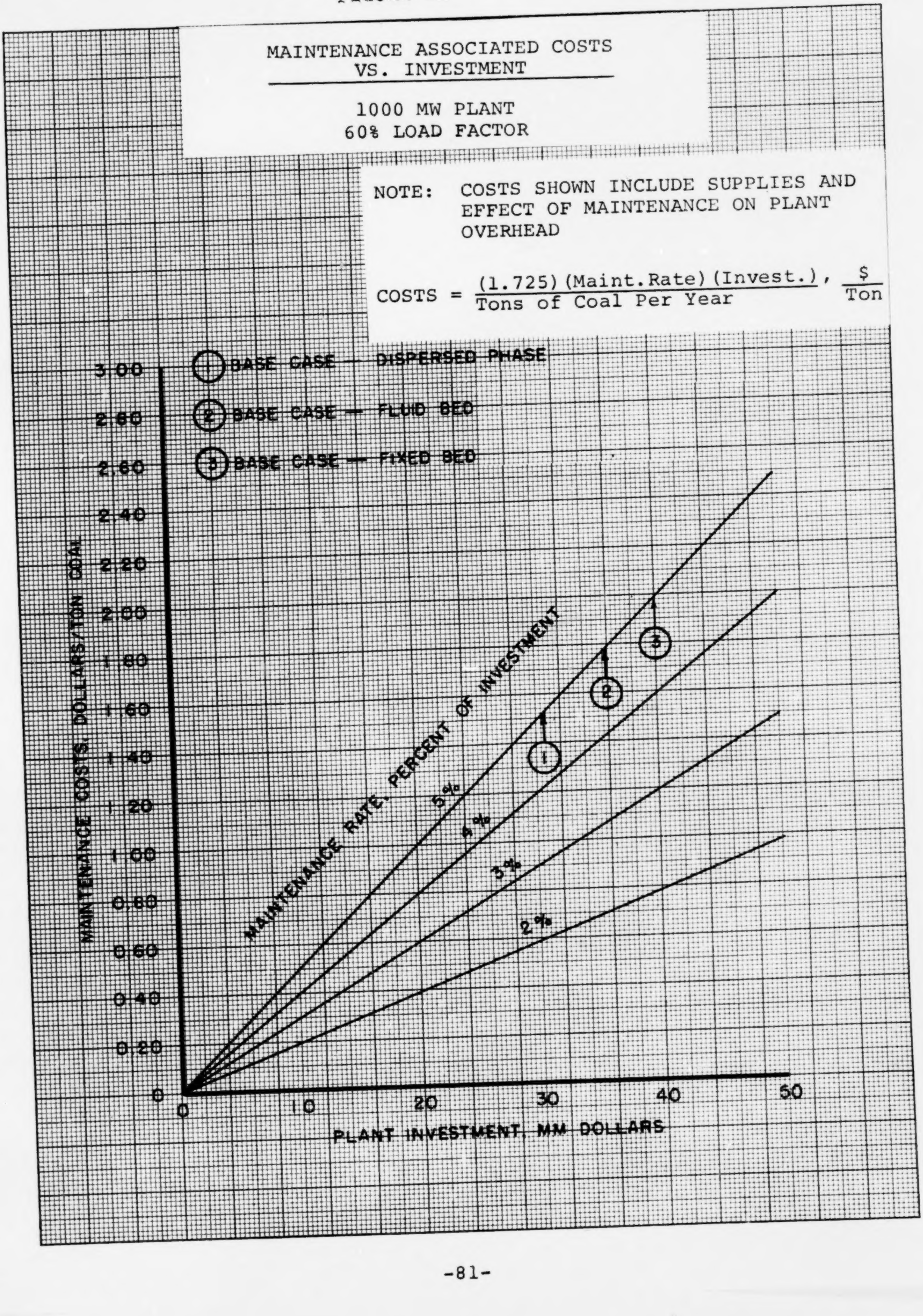
2. MAINTENANCE COSTS

Since in the present study maintenance costs are calculated as a percentage of plant investment the total annual cost attributable to maintenance varies with the charge rate but is unaffected by load factor. On the other hand, the unit costs for any given charge rate do vary with load factor. Thus in order to calculate maintenance costs it is necessary to specify both charge rate and load factor. A means of evaluating a range of each is presented below.

All of the alkalized alumina plant costs shown in this report include annual maintenance costs equal to 5% of fixed capital investment; the 5% figure is the sum of the assumed labor (3%) and material (2%) costs. The individual labor and material costs may differ from that assumed but the overall value of 5% is felt to be a reasonable basis for calculating the base case costs. Past experience shows that the annual maintenance costs of a typical refinery are about 4% and those of a low overhead chemical plant average about 4.5% for the first years of operation. After 10 years the maintenance costs are higher. Since the cited refinery and chemical plant represent well established processes as opposed to the commercially unproven alkalized alumina process, the use of slightly higher maintenance costs appears reasonable. It is possible, however, that maintenance costs might differ from the values assumed, and therefore it is of interest to determine their effect on plant costs for a range of values. Because of the high overall base case plant costs it was decided to limit the evaluation of maintenance costs to values less than used in the base case. Consequently annual costs associated with maintenance charge rates ranging from 2% to 5% have been calculated for plant investments up to $50,000,000, and the results are shown in Figure 11 for a 60% load factor. The effect of load factor on maintenance costs is shown in Figure 12.

The costs shown in Figures 11 and 12 include all of the maintenance associated charges. That is, in the present evaluations a method of calculating costs is used whereby supplies and plant overhead charges are calculated as functions of maintenance costs and these effects have been included in the Figures 11 and 12 values. Since supplies are assumed to be 15% of maintenance, and plant overhead includes 50% of maintenance and supplies, the total contribution of maintenance to plant cost is (1.725) X (maintenance rate) X (investment). In Figures

FIGURE 11



MAINTENANCE ASSOCIATED COSTS VS. INVESTMENT

1000 MW PLANT
60% LOAD FACTOR

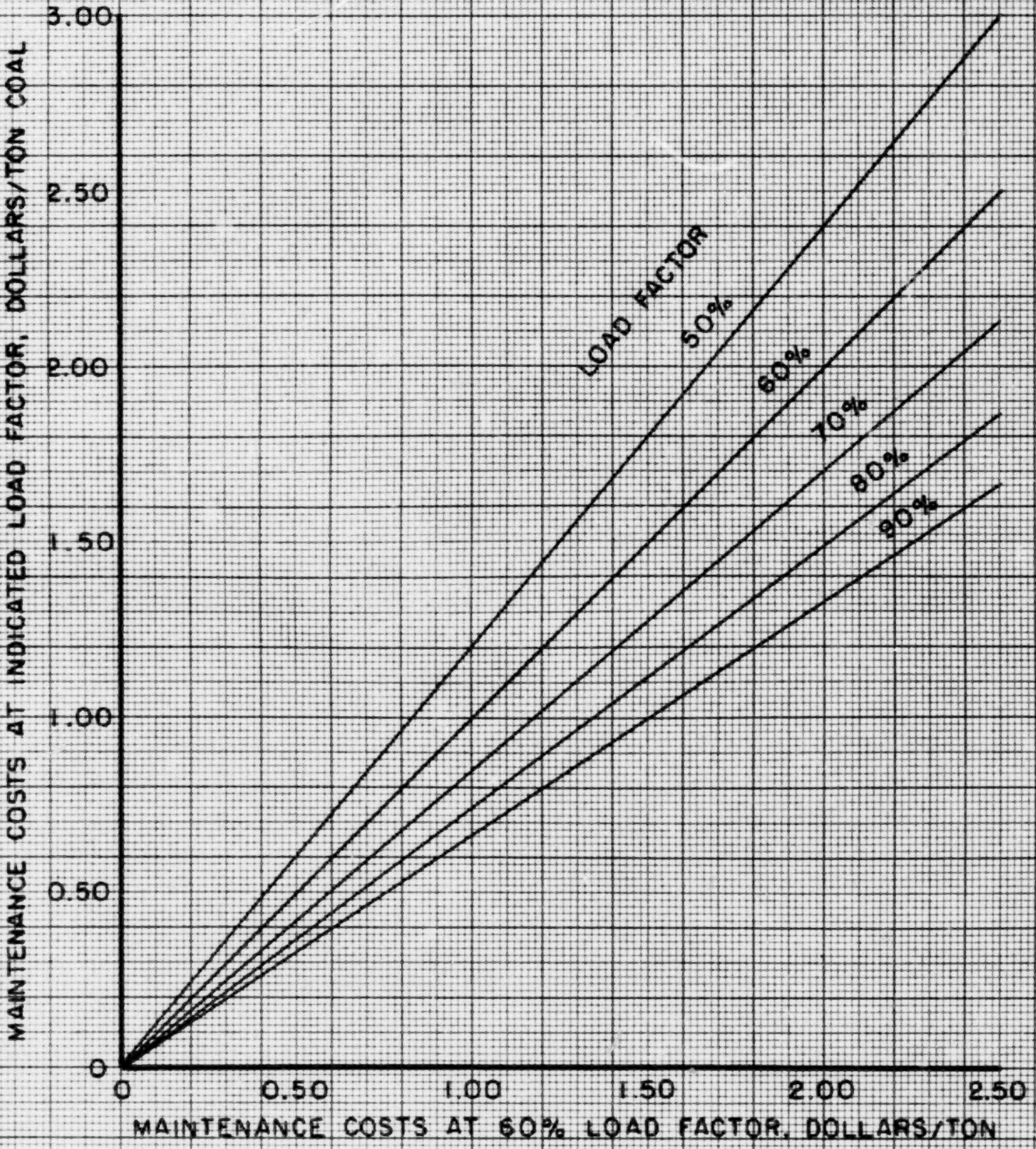
NOTE: COSTS SHOWN INCLUDE SUPPLIES AND EFFECT OF MAINTENANCE ON PLANT OVERHEAD

$$\text{COSTS} = \frac{(1.725)(\text{Maint.Rate})(\text{Invest.})}{\text{Tons of Coal Per Year}}, \ \frac{\$}{\text{Ton}}$$

FIGURE 12

EFFECT OF LOAD FACTOR ON MAINTENANCE ASSOCIATED COSTS
1000 MW PLANT

PERCENT LOAD FACTOR	HOURS PER YEAR
50	4375
60	5250
70	6125
80	7000
90	7875



11 and 12 the costs thus calculated are reported using the previously established unit operating cost standard; i.e., dollars per ton of power plant coal burned.

The maintenance costs for the three base case designs are indicated on Figure 11 and are 1.53, 1.78 and 2.00 dollars per ton of coal for the dispersed phase, fluid bed, and fixed bed respectively. The change in plant cost resulting from use of different maintenance charges is readily obtained from Figures 11 and 12 as follows: for any specified plant investment, load factor and maintenance charge rate, read maintenance costs in dollars per ton of coal from Figures 11 and 12; from this value subtract the base case value given above and the difference is the change in plant cost. For example, determine the plant cost for a dispersed phase plant (investment = $30,720,000) operating at 80% load factor and with a maintenance charge rate of 4%. From Figure 11 maintenance costs at 60% load factor and 4% are $1.22/ton. From Figure 12, costs at 80% load factor are $0.92/ton coal. Thus the change in plant operating cost is (0.92 -1.53) or -$0.61/ton coal. Since the dispersed phase base case cost is $4.53/ton, the new cost is 4.53 -0.61 or $3.92/ton coal. Thus increasing the load factor to 80% and lowering maintenance costs to 4% results in a net savings of $0.61/ton for the dispersed phase design. Other combinations can be obtained in a similar manner.

It is of interest to note the change in total plant costs as the maintenance charge rate is decreased while keeping all the other variables constant. From Figure 11 the cost change resulting from lowering the maintenance charge rate has been determined for each process design and the results are summarized below.

MAINTENANCE CHARGE RATE		4%	3%	2%
	BASE CASE COST, $/TON	PERCENT DECREASE IN COST		
DISPERSED PHASE	4.53	6.8	13.5	20.3
FLUID BED	4.98	7.0	14.2	21.5
FIXED BED	5.52	7.3	14.5	21.8

The above results indicate that a 1% decrease in maintenance charge rate lowers the plant costs by about 7%. Thus at annual maintenance rates of 4%, reductions in plant costs of only about 30-40 cents per ton are realized. The 3% and 2% maintenance charge rates are shown to demonstrate their effect on plant costs but it is not expected that these low rates would normally be encountered in an actual plant.

D. UTILITIES AND CREDITS

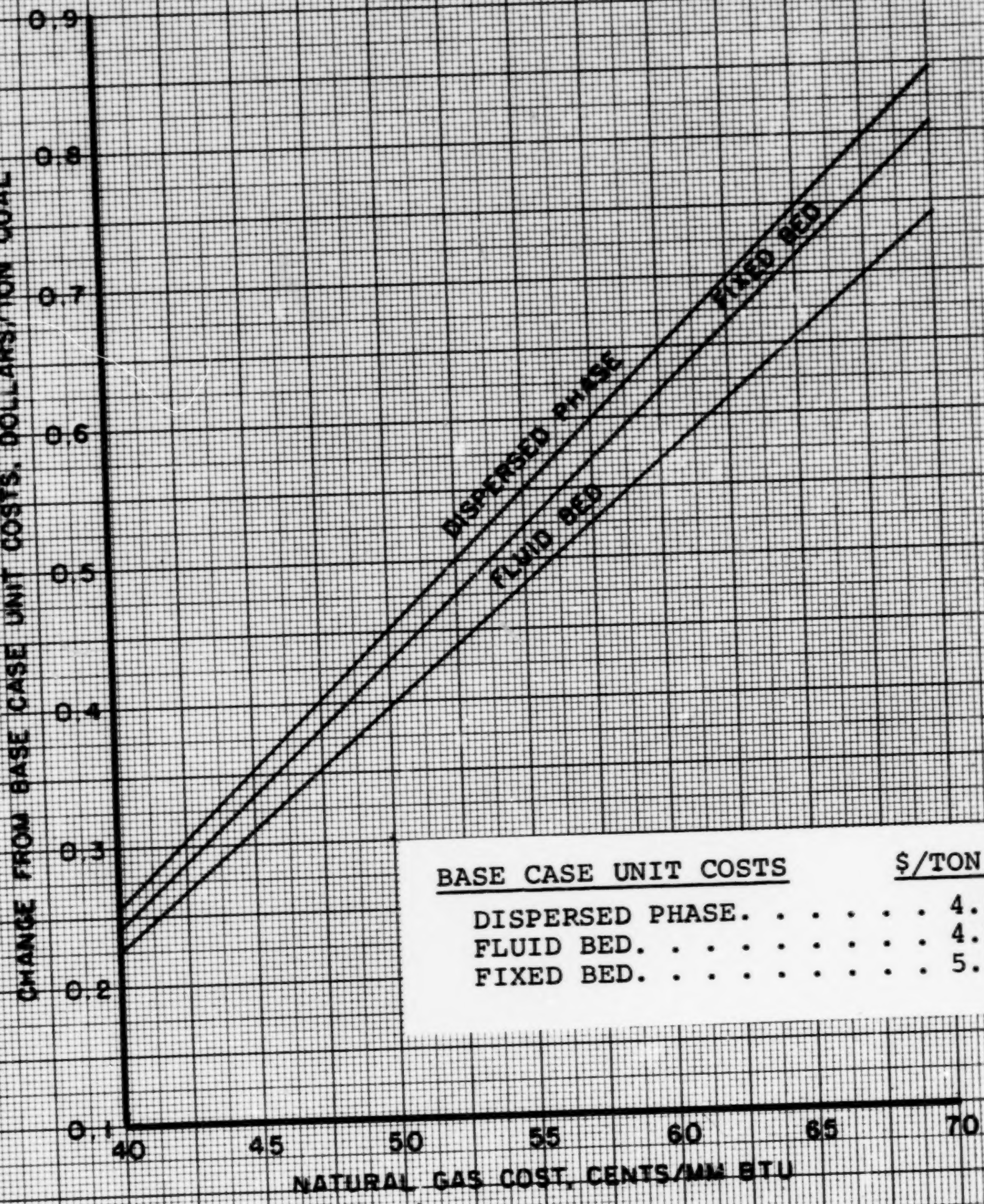
1. NATURAL GAS COST

Natural gas is required in the SO_2-removal plant for two different operations: 1) as feed to the steam-methane reformer which produces the sorbent regeneration gas; and 2) as fuel for the solids heater wherein the sorbent is heated from the base case sorption temperature of 600°F to the regeneration temperature of about 1400°F. Other sources of reducing gas could be used (e.g., producer gas) if natural gas were not available but based on published values, it appears that reformed gas offers the lowest overall costs. Likewise, other types of fuel could be used for heating the sorbent to its regeneration temperature but again the overall costs would increase. Therefore, it was decided to use natural gas for the present evaluations with the reservation that it may not be available for some specific sites or, if available, the costs might be exorbitant. Accordingly, a range of natural gas prices has been evaluated for their effect on the cost of SO_2-removal using the consumption rates calculated for the base case designs.

The effect of natural gas price on alkalized alumina plant costs is shown in Figure 13 The base case economics are based on a price of 27 cents per million Btu (¢/MM Btu), which is the average price of natural gas used by power plants in 1967. (Steam Electric Plant Factors/1968 Edition, Natural Coal Association, Washington, D.C.) It should be noted, however, that this figure represents what is essentially the "summer" price, and a guaranteed year-round supply of natural gas would most likely command a significantly higher price, probably in the order of 50¢/MM Btu. Figure 13 shows the effect of natural gas price up to 70¢/MM Btu which should cover the range that

FIGURE 13

EFFECT OF NATURAL GAS COST ON BASE CASE OPERATING COSTS 1000 MW POWER PLANT



BASE CASE UNIT COSTS	$/TON COAL
DISPERSED PHASE.	4.53
FLUID BED.	4.98
FIXED BED.	5.52

will apply to power plants. If costs outside those shown in Figure 13 are desired, the curves can be readily extrapolated since the plant cost/natural gas price relationship is linear. Assuming a natural gas price of 50¢/MM Btu does apply, Figure 13 shows that the increased cost to the base case is \$0.40/ton for the fluid bed, \$0.43/ton for the fixed bed, and \$0.45/ton for the dispersed phase. Thus raising the natural gas price to what is probably a more realistic value than used in the base case costs increases these costs by 8-10%.

2. STEAM CREDIT

Operation of the plants produces steam in excess of that required by the process and credit has been taken for this steam. It is assumed that the excess steam could be used either by the power plant itself or to drive the fans and compressors in the SO_2-removal plant; in either case the value of this steam can be used to reduce costs. The value of excess steam, assuming there is a use for it, depends largely on the local price of fuel as well as on the quality (pressure, temperature). For the present study the value of the low pressure (50 psig) steam produced in the SO_2-removal plant is assumed to be equal to the fuel value. The base case costs therefore include a steam credit based on a unit cost of 25 cents per thousand pounds (¢/M lbs). Since the contribution of the excess credit will vary depending on the unit cost of the steam, a range of values has been evaluated for their effect on plant costs for all three process designs and the results are shown in Figure 14. As can be seen from Figure 14 the value assigned to excess steam has very little effect on SO_2-removal operating costs. For example, increasing the steam value from 25 to 50¢/M lbs reduces the unit costs by about $0.06/ton coal, or less than 2%. It should be noted that the effect of steam credit on plant costs is independent of load factor. Also, if steam costs outside the range shown are desired the curves may be readily extrapolated since the relationship between unit costs and steam credit is linear.

3. SULFUR CREDIT

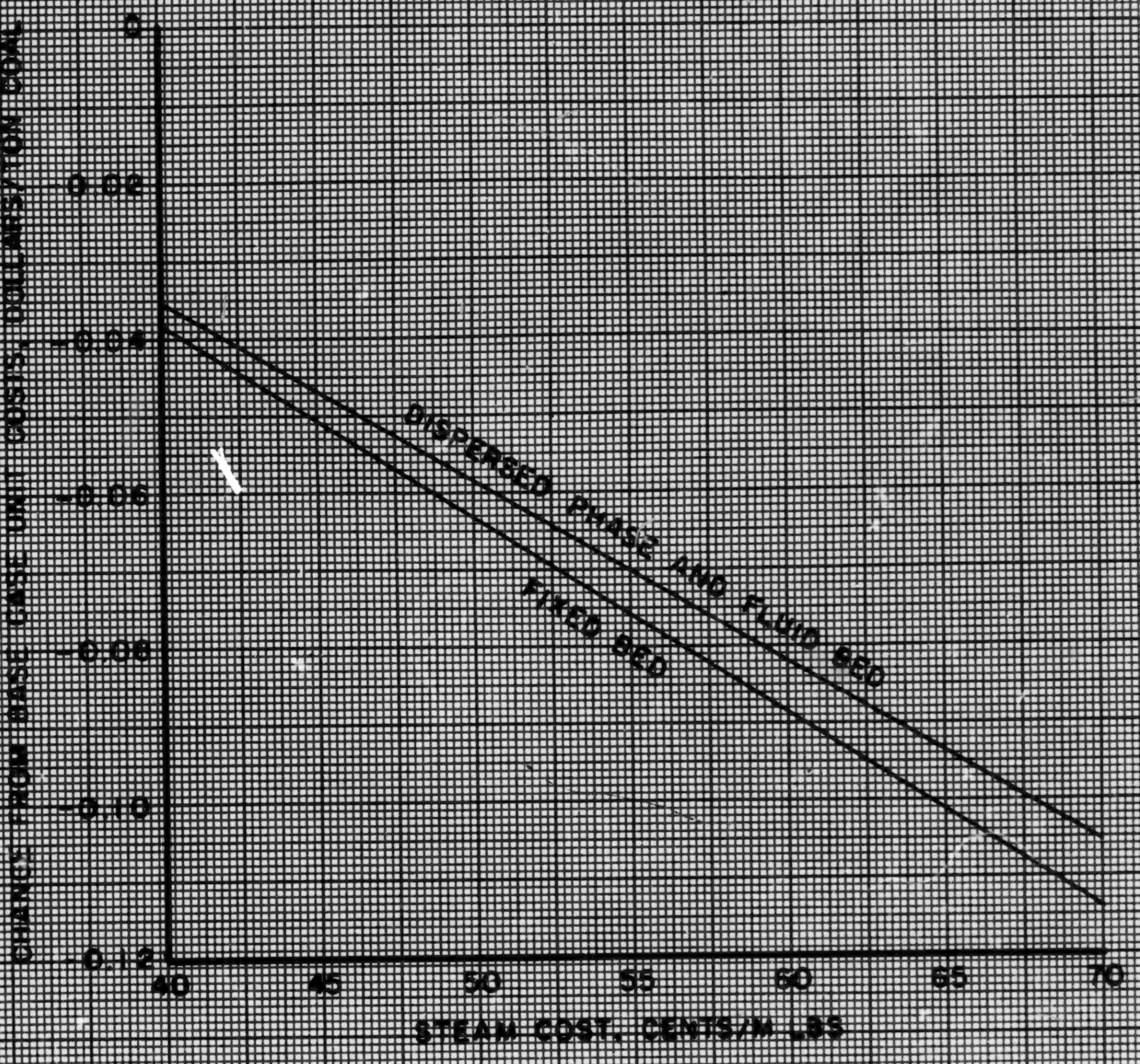
The SO_2 removal from the flue gas is recovered eventually as elemental sulfur which can be sold, thus offsetting at least part of the SO_2-removal plant operating costs. The total revenue obtained from the sale of by-product sulfur will, of course, be dependent upon the quantity recovered and the selling price. For any specific set of conditions (e.g., 3% sulfur coal, 90% recovery) the sulfur credit will be constant for each selling price when expressed as unit costs, i.e., $/ton coal.

The sulfur credit for the base case conditions for each process is shown on page 94. Four levels of sulfur price have been evaluated, viz., $0, $20, $30, and

FIGURE 14

EFFECT OF STEAM CREDIT ON OPERATING COSTS

1000 MW POWER PLANT



$40/short ton of sulfur and the effect on SO_2-removal plant operating costs determined. The net reduction of unit costs attributable to the by-product sulfur ranges from 0 to $1.08/ton coal for the dispersed phase and fluid bed designs, and from 0 to $1.16/ton for the fixed bed case. The increased credit for the latter design is owing to the higher sulfur recovery obtained with the fixed bed process. The base case credits are based on a $20/ton sulfur price which was selected as described in the phases 1 and 2 final report and therefore will not be repeated here.

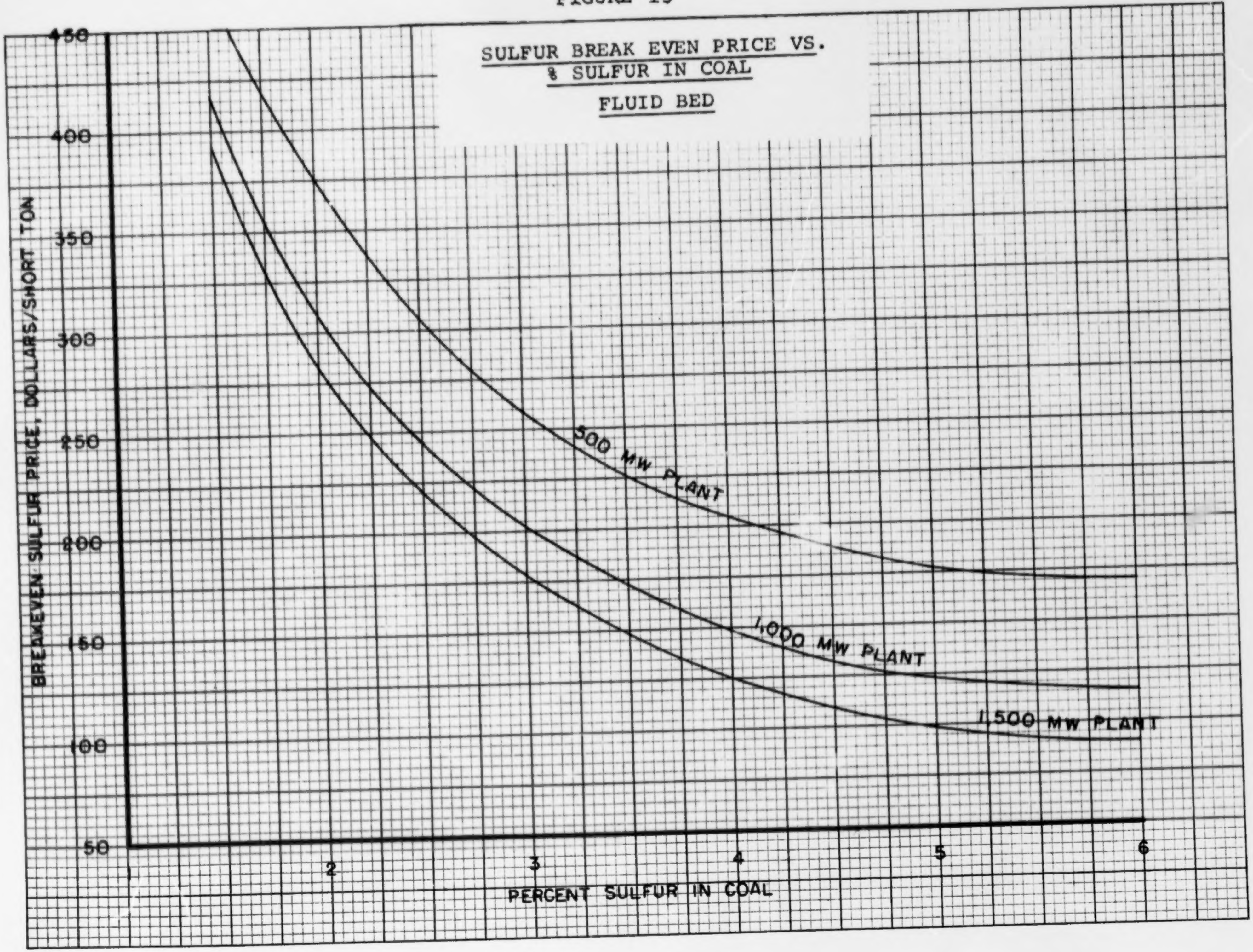
The effect of by-product sulfur on unit costs is, of course, 0 when the selling price is 0 and while this situation probably will not occur, it is a possibility. That is, it is conceivable that local conditions could be encountered whereby there would be no nearby market for the sulfur and disposal would be a problem. It is not anticipated, however, that negative costs would ever be encountered; i.e., it is assumed that storage, handling, and shipping costs will always be recovered.

The effect on operating costs of burning higher and lower sulfur coals was determined (see Section E below) and the results show that the costs increase as the amount of sulfur in the coal increases. It might be expected that as the sulfur content of the coal increases, hence producing more by-product credit, the unit costs would decrease whereas in fact they increase. Even though more sulfur is recovered from the higher sulfur coals, the selling price is less than production costs so the net cost of the SO_2-removal plant increases.

To determine if there is a possibility of making a profit from the operation of a SO_2-removal plant, the sulfur break even price has been calculated; i.e., the price at which sulfur would have to be sold to recover the total annual operating cost of the SO_2-removal plants. The results are plotted in Figures 15 (fluid bed) and 16 (dispersed phase) for a 1000 MW plant. Also indicated on Figures 15 and 16 are break even prices for a 500 MW and 1500 MW plant. These latter two cases are extrapolated since only the values at 3% sulfur are known. However, it is expected that the curves would be about parallel as shown and the results are therefore accurate enough to draw valid conclusions. Based on these results it is apparent that to obtain a zero cost for the SO_2-removal plant, sulfur would have to be sold for prices far exceeding those that can be reasonably expected. For example, with a 1500 MW plant and 6% sulfur coal, a sulfur price of about $90/ton is required for the fluid bed process and this rises to about $170/ton when the plant size is reduced to 500 MW. For plants using 3% sulfur coal, selling prices in the range of $175-250/ton are required. The results in Figures 15 and 16 show also that the break even selling price levels off in the range of 6% sulfur and use of higher concentrations (e.g., refuse from a coal cleaning operation) would not appreciably reduce costs.

Since the sulfur break even selling price discussed above is so much higher than the actual selling price that reasonably can be expected, another break even price was calculated using a different concept as follows. The lowest break even price possible will be obtained at the minimum operating cost which in turn results when the least amount of equipment is used. Minimum equipment is a function of the process and for a chemically regenerable process a source of reducing gas is required and a Claus plant is needed to produce the sulfur. Thus these two items, i.e., reformer and Claus plant, represent the absolute minimum quantity of equipment needed to regenerate a sorbent and produce sulfur. It was assumed, therefore, that except for the

FIGURE 15



SULFUR BREAK EVEN PRICE VS.
% SULFUR IN COAL

FLUID BED

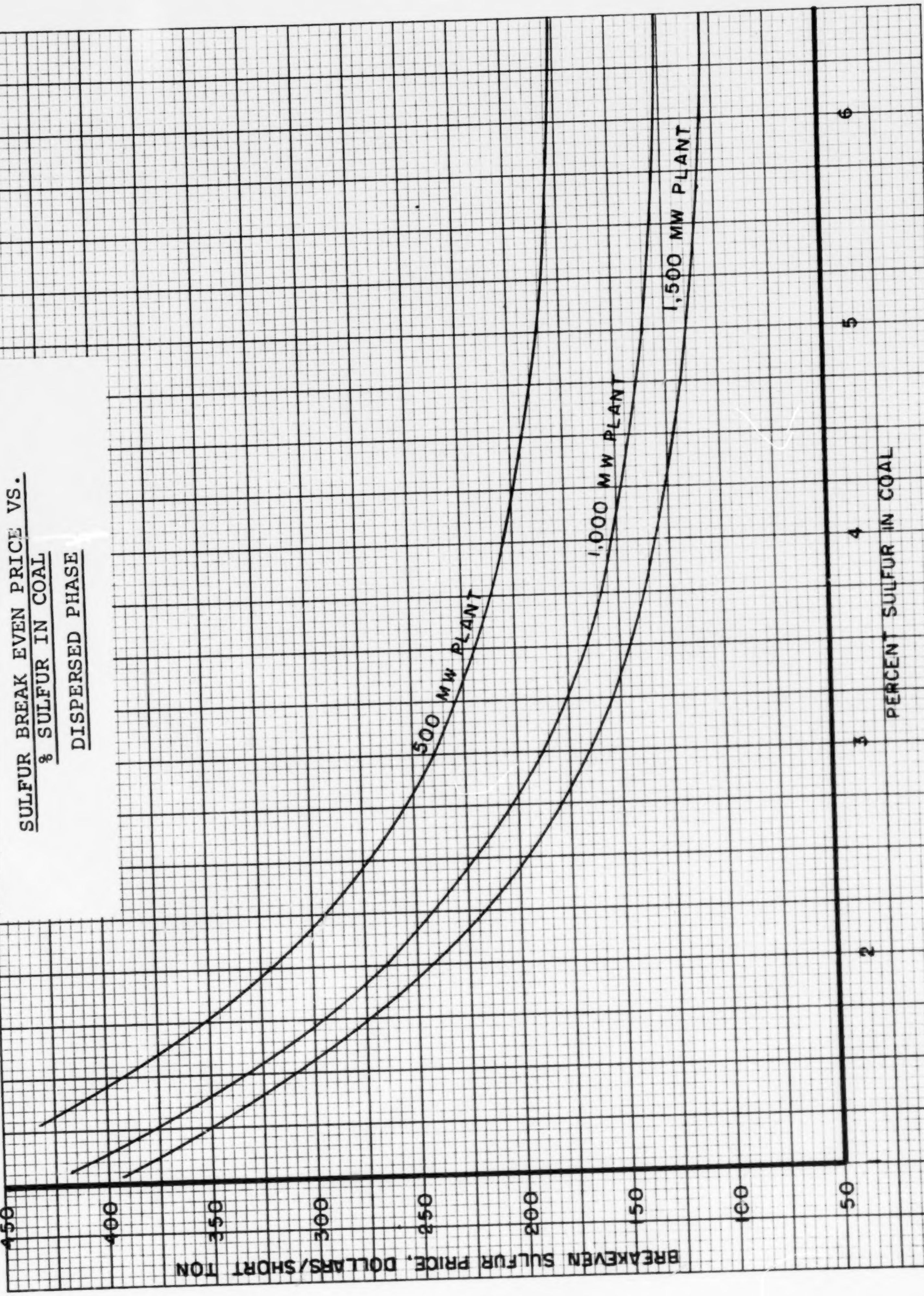


FIGURE 16

SULFUR BREAK EVEN PRICE VS. % SULFUR IN COAL

DISPERSED PHASE

reformer and Claus plant, installation and operation of the SO_2-removal plant could be achieved at zero cost.

Using the concept just described the break even selling price of sulfur was calculated for the base case conditions and a value of \$36.84/short ton was obtained for the fluid bed and dispersed phase designs whereas the fixed bed process requires a price of \$53.79/short ton. Since the actual selling price is expected to be \$20/ton or less, it is obvious that a net loss would result for all sulfur produced even if there are no charges for the rest of the SO_2-removal plant. Based on these results it is concluded that sale of by-product sulfur cannot offset the cost of the SO_2-removal plant.

To include sulfur credit in the SO_2-removal plant operating cost deduct the Δ cost shown in the table below. These costs are based on a 3% sulfur coal but credits for other concentrations can be obtained by direct proration since the difference in coal heating value and/or power plant heat rate will have only a small effect on quantity of sulfur recovered. Thus a 6% sulfur coal will have credits that are twice those shown below.

SULFUR PRICE	SULFUR CREDIT, Δ COST, $/TON COAL		
$/TON	DISPERSED PHASE	FLUID BED	FIXED BED
20	0.54	0.54	0.58
30	0.81	0.81	0.87
40	1.08	1.08	1.16

E. POWER PLANT OPERATION

1. SULFUR CONTENT OF COAL

The sulfur content of the coal burned in the power plant for the base case evaluations is 3%, producing an SO_2 concentration in the flue gas of 2146 ppm of which 90% is removed, yielding a treated flue gas containing 215 ppm SO_2. To determine the effect of sulfur concentration in the coal on SO_2-removal plant costs, two additional cases have been developed for both the dispersed phase and fluid bed processes, viz, designs based on burning coals containing 1-1/2% S and 6% S. In both of the latter designs the SO_2 concentration in the exit gas has been kept at the 215 ppm level used in the base cases. This approach assumes, of course, that pollution control laws will be based on exit gas concentration rather than degree of removal of entering sulfur but at the present time this seems to be a reasonable assumption.

It was originally planned to evaluate sulfur concentrations up to 12% which could be obtained in the refuse from a coal beneficiation plant designed to produce low sulfur coal. As such, the ash content would be very high, 35-50%, and the heating value (7630 Btu/lb as per calculations summarized in Table 21). Because of the low heating value and high ash content, which is some 3 to 5-fold higher than normally found in power plant coals, it was decided that the 12% S coal does not represent a realistic case and therefore was not evaluated. It is doubtful that a dependable supply of this coal could be obtained in the quantity needed by a large size power plant, and if it could be obtained, ash disposal costs probably would be prohibitive. Consequently, evaluations of the sulfur content of coal were restricted to a maximum of 6%.

The economic effects of the coal sulfur concentration have been determined as changes from the base case values. The investment changes are shown as a function of the sulfur content of the coal in Figure 17 for both the dispersed phase and fluid bed designs. It is apparent from Figure 17 that the SO_2-removal plant investment changes significantly as the sulfur content of the coal varies between 1-1/2% S and 6% S with the fluid bed design being less affected than the dispersed phase at higher sulfur levels. For example, at 1-1/2% S the dispersed phase investment is reduced by about $4.7 million (15%) compared to a $4.1 million (11%) reduction for the fluid bed while the corresponding values at 6% S are increases of $9 million (29%) and $5 million (14%). The changes in investment shown here were derived from the base case estimates prepared for the phases 1 and 2 studies.

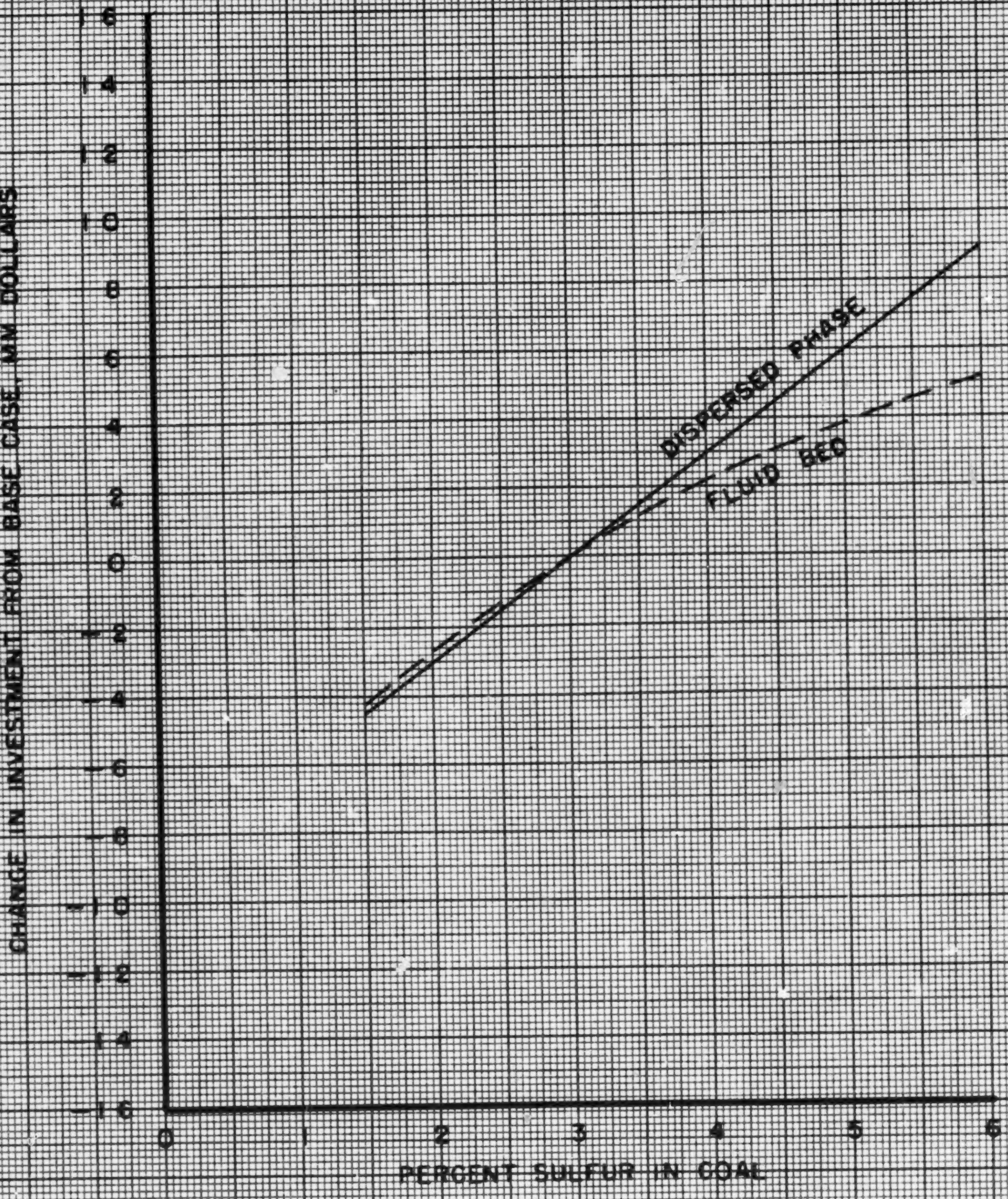
The larger increase in process plant investment obtained for the dispersed phase design than for the fluid bed is owing chiefly to the smaller amount of sulfur removed per unit weight of sorbent which in turn results in much higher solids circulating rates. The larger solids circulating rate causes increases in the sizes and cost of all of the dispersed phase process equipment (reactor, regenerator, etc.), comprising about 45% of the total incremental investment. The remaining 55% arises from increases in the cost of the Claus plant, reformer, electrostatic precipitator, and air preheater. The incremental costs of these latter four items are nearly the same in both designs but part of the fluid bed process equipment (e.g., reactor) remains the same. Further, because of the nature of the design (longer reaction contact time, higher sulfur loadings, less solids circulation, etc.), the equipment that does change requires less modification than does that of the dispersed phase plant.

FIGURE 17

CHANGE IN INVESTMENT FROM BASE CASE
VS. SULFUR CONTENT OF COAL
1000 MW PLANT

BASE CASE INVESTMENTS

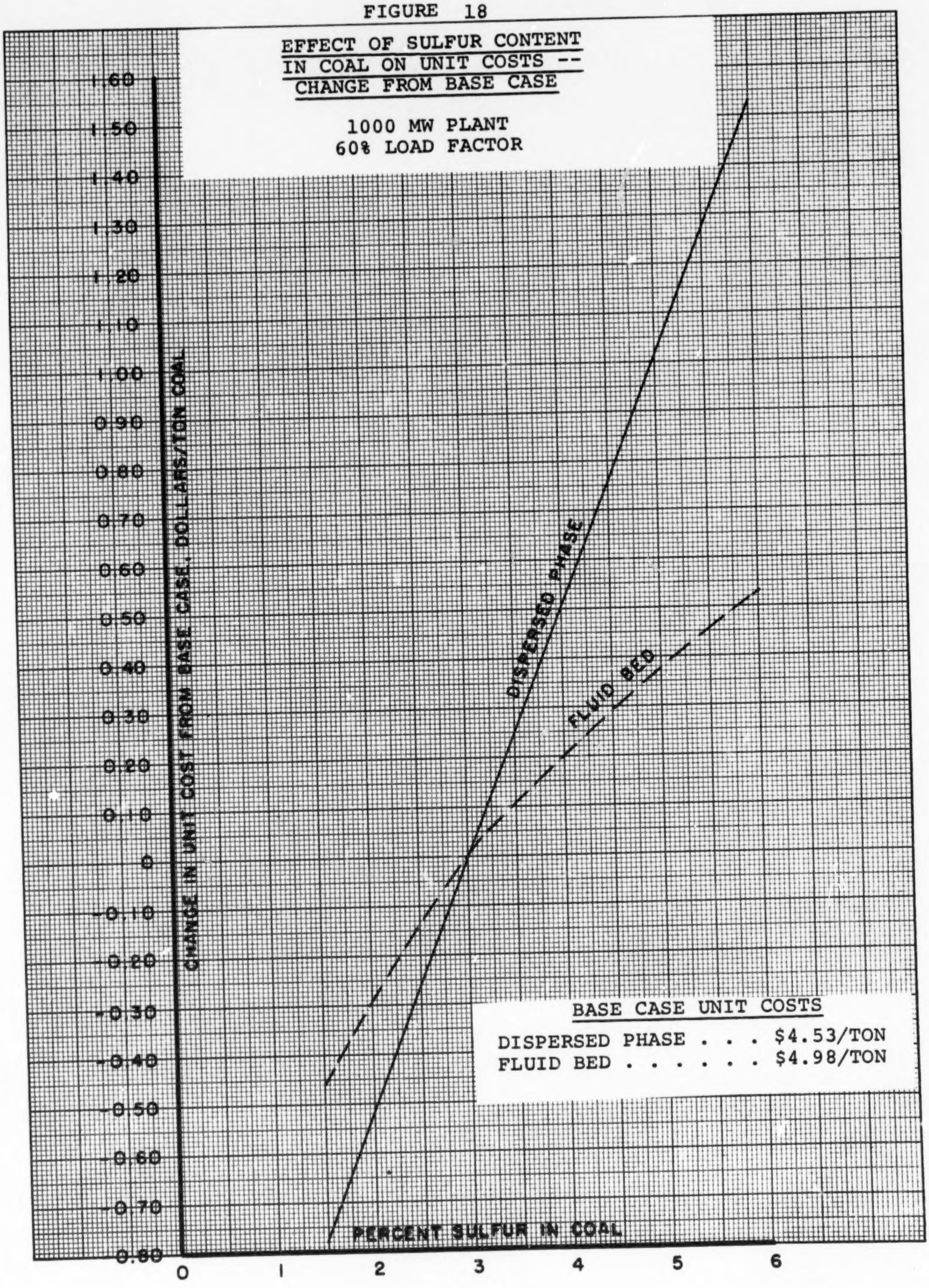
DISPERSED PHASE.	$30,720,000
FLUID BED.	$35,900,000



Since the hot regenerated solids are sent directly from the regenerator into the reactor, the final mixed (i.e., equilibrium) temperature in the reactor is a function of the quantity of solids as long as the other parameters (e.g., flue gas volume, heat losses, etc.) are essentially constant - as is the actual case -. With the greater solids flow in the dispersed phase design, when the higher sulfur coals are used, the temperature rises, hence flue gas volume, is higher than for the fluid bed. This is the primary reason for the change in reactor size and cost. On the other hand, the design of the base case fluid bed reactors is determined more by mechanical and operability considerations than by process requirements, and the design changes owing to changes in sulfur concentration in the coal are not large enough to affect the basic design. Therefore, while the dispersed phase plant shows a linear relationship between sulfur in coal and investment, the fluid bed investment increases at a much lower rate at sulfur levels above 3%. In fact, for coal having 6% sulfur, the fluid bed plant investment approaches that of the dispersed phase, being only about $1.3 million more whereas the difference is about $5 million for coals containing 3% sulfur and less.

The change in unit operating costs from the base case (dollars/ton coal) for both the dispersed phase and fluid bed designs are shown in Figure 18 as a function of the sulfur concentration in the coal. The general shapes of the curves are the same as the investment curves shown in Figure 17; e.g., the dispersed phase costs show a linear relationship with percent sulfur whereas the rate of change in fluid bed costs decreases with sulfur content. In fact, for a 6% sulfur coal the fluid bed unit costs are less than the dispersed phase (5.51 vs. 6.04 dollars per ton) even though the investment is $1.3 million higher

FIGURE 18



EFFECT OF SULFUR CONTENT IN COAL ON UNIT COSTS -- CHANGE FROM BASE CASE

1000 MW PLANT
60% LOAD FACTOR

BASE CASE UNIT COSTS

DISPERSED PHASE . . . $4.53/TON
FLUID BED $4.98/TON

(41.0 vs. 39.7). This reversal stems from the higher attrition losses in the dispersed phase design resulting from the increased circulating solids rate. As is evident from Figure 18, the unit costs increase with sulfur content in both cases.

The results of the coal sulfur study are summarized below:

% SULFUR IN COAL	INVESTMENT, M $		OPERATING COST, $/TON COAL	
	CHANGE	NEW	CHANGE	NEW
FLUID BED:				
1-1/2	-4,270	31,630	-0.46	4.52
3 (Base)	0	35,900	0	4.98
6	+5,060	40,960	+0.53	5.51
DISPERSED PHASE:				
1-1/2	-4,670	26,050	-0.79	3.74
3	0	30,720	0	4.53
6	+8,930	39,650	+1.51	6.04

2. FLUE GAS VOLUME

To determine the effect of flue gas volume on plant costs, flue gas production was calculated as a function of the following variables over the ranges indicated:

(a) Sulfur content of coal -- 1.5 to 12 wt % S
(b) Power plant size -- 500 to 1500 MW
(c) Heating value of coal -- 10,000 and 13,000 Btu/lb
(d) Power plant heat -- 8600 and 9000 Btu/kwh

Flue gas quantities were calculated for 36 different combinations of the above variables and the results are summzrized in Tab le 21. Flue gas volumes are not shown since the temperature and pressure varies from case to case.

TABLE 21

FLUE GAS QUANTITY AS A FUNCTION OF PLANT SIZE, HEAT RATE, SULFUR CONTENT, AND HEATING VALUES OF COAL

PLANT SIZE, MW	BTU/LB COAL	BTU/KWH	COAL BURNED, LB/HR	FLUE GAS PROD., MOLES/HR
BASE CASE (3 WT % SULFUR)				
500	13,000	8600	330,769	144,291
500	"	9000	346,154	151,003
1000	"	8600	661,538	288,583 *
1000	"	9000	692,308	302,006
1500	"	8600	992,308	432,875
1500	"	9000	1,038,462	453,008
CASE I (2.44 WT % SULFUR)				
500	10,000	8600	430,000	152,220
500	"	9000	450,000	159,300
1000	"	8600	860,000	304,440
1000	"	9000	900,000	318,600
1500	"	8600	1,290,000	456,660
1500	"	9000	1,350,000	477,900
CASE II (1.5 WT % SULFUR)				
500	13,160	8600	326,748	144,060
500	"	9000	341,945	150,760
1000	"	8600	653,495	288,119
1000	"	9000	683,891	301,521
1500	"	8600	980,243	432,179
1500	"	9000	1,025,836	452,281
CASE III (6 WT % SULFUR)				
500	12,700	8600	338,583	144,676
500	"	9000	354,331	151,406
1000	"	8600	677,165	289,353
1000	"	9000	708,661	302,811
1500	"	8600	1,015,748	434,029
1500	"	9000	1,062,992	454,216
CASE IV (12 WT % SULFUR)				
500	7,630	8600	563,565	172,248
500	"	9000	589,777	180,259
1000	"	8600	1,127,130	344,496
1000	"	9000	1,179,554	360,519
1500	"	8600	1,690,695	516,744
1500	"	9000	1,769,332	540,779

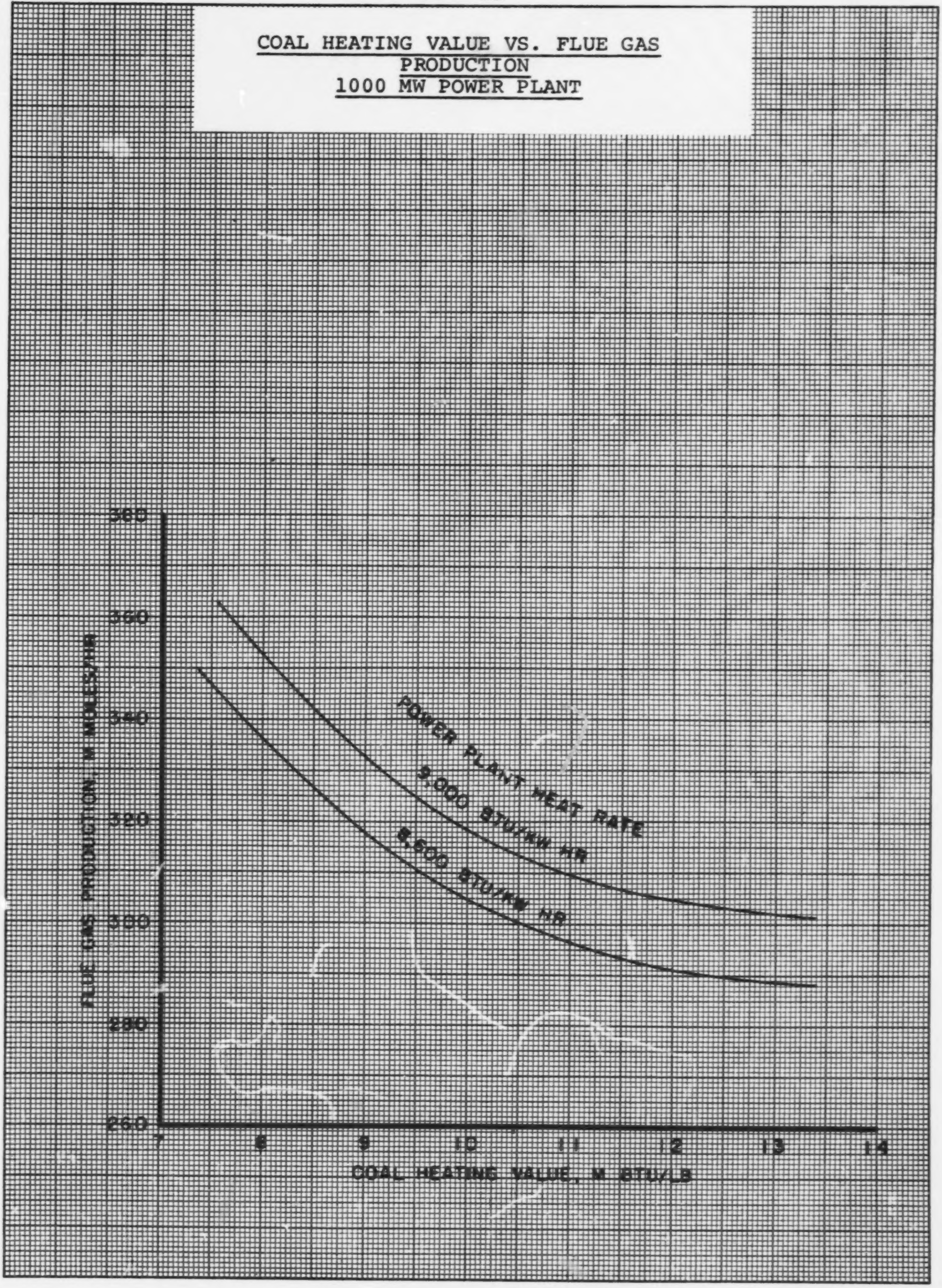
* **ORIGINAL BASE CASE**

The original process designs are based on treating the flue gases from a 1000 MW power plant burning a 3 wt % sulfur coal, having a heating value of 13,000 Btu/lb, at a heat rate of 8600 Btu/kwh. In the base case calculations using a coal heating value of 10,000 Btu/lb, the sulfur content was adjusted from 3.0 wt % to 2.44 wt % so that the flue gas mole fraction of SO_2 would remain unchanged (i.e., 2150 ppm) and thus permit the effect of flue gas volume on economics to be isolated. The coal analyses for the different cases are derived from the base case and adjusted to give the desired heating value and sulfur content. It should be noted that nearly all of the coals presently used in power plants have heating values in the range 10,000 - 13,000 Btu/lb, which is the reason these values were selected for the present study.

Comparing Case I with the base case shows that a reduction in coal heating value from 13,000 to 10,000 Btu/lb results in an increase of 5.5% in flue gas volume. A similar comparison shows that raising the power plant heat rate from 8600 to 9000 Btu/kwh causes an increase of 4.7% in flue gas volume. Referring to Figure 19, which is a plot of coal heating value vs. flue gas volume, it can be seen that the 5.5% represents an average overall change for the range of 10,000 to 13,000 Btu/lb. Since the plot is not a straight line function special cases represented by the 12% sulfur coal would require corrections up to 20%; however, this case was eliminated as discussed in Section E-1.

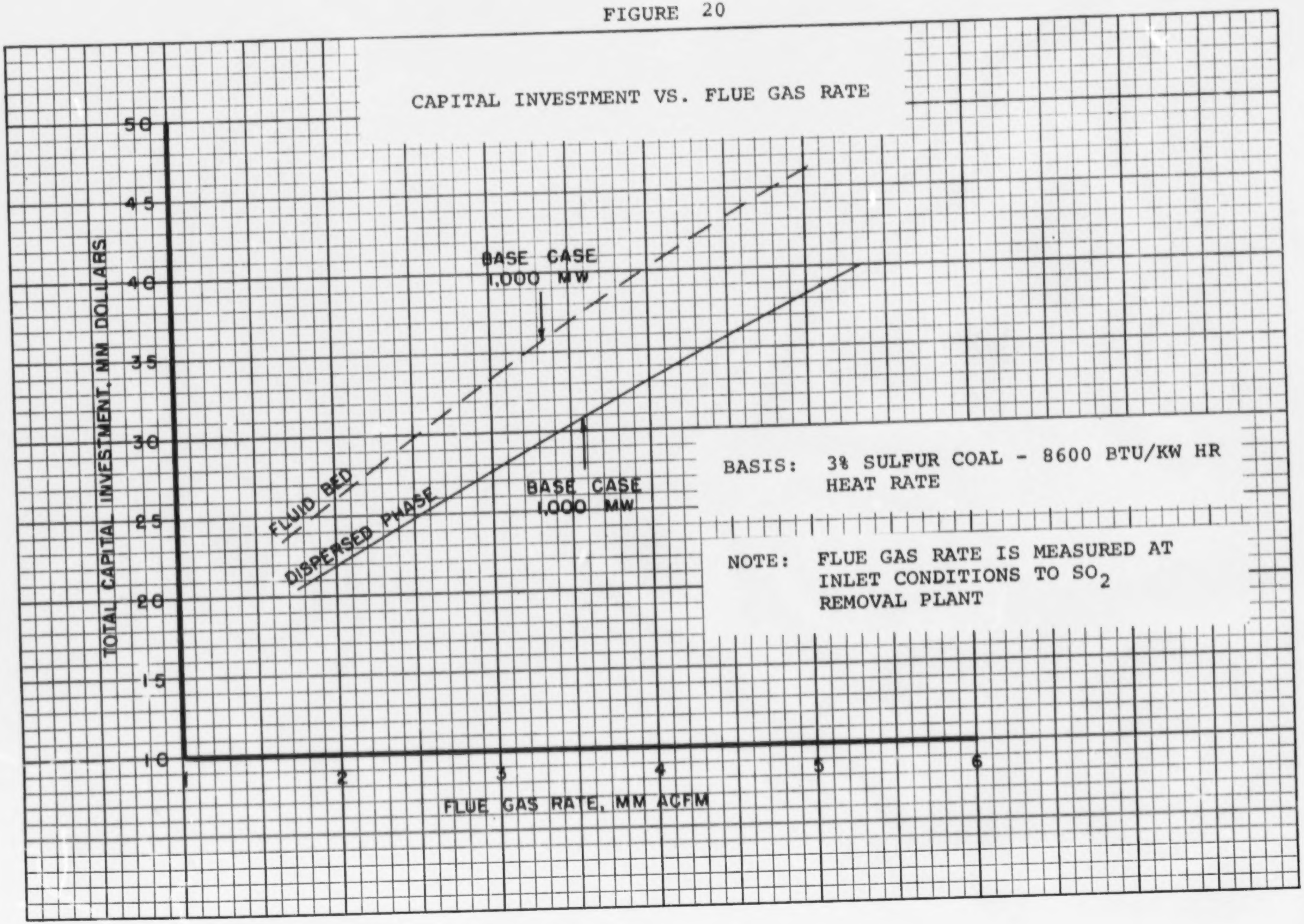
Plots of plant investment and operating costs as a function of flue gas flow rate are shown in Figures 20 and 21. Thus, the flue gas volumetric flow rate for any desired conditions in the range evaluated can be calculated from the data in Table 21 using the appropriate temperature and pressure, and the corresponding investment

FIGURE 19



COAL HEATING VALUE VS. FLUE GAS PRODUCTION
1000 MW POWER PLANT

FIGURE 20

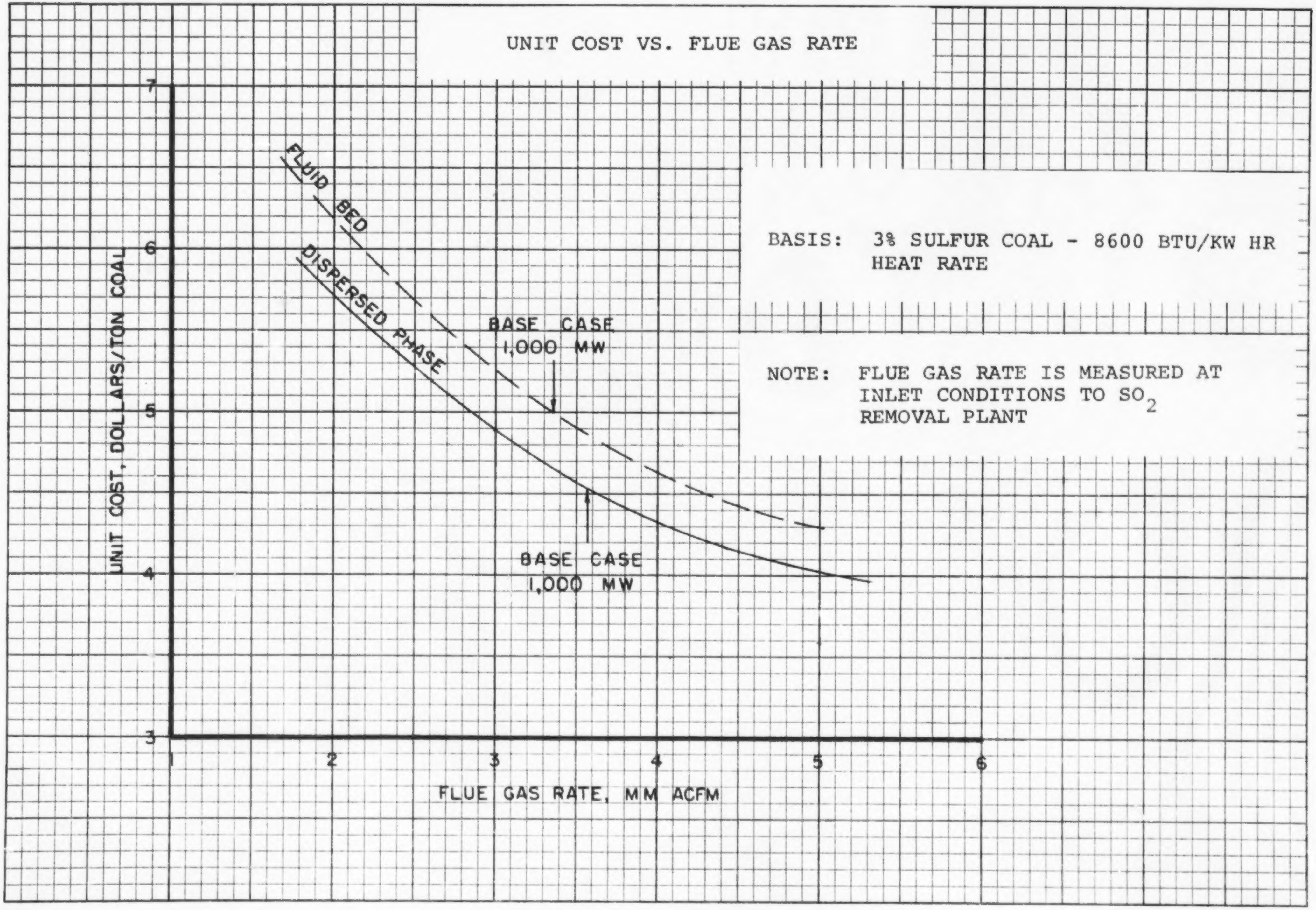


CAPITAL INVESTMENT VS. FLUE GAS RATE

BASIS: 3% SULFUR COAL - 8600 BTU/KW HR HEAT RATE

NOTE: FLUE GAS RATE IS MEASURED AT INLET CONDITIONS TO SO_2 REMOVAL PLANT

FIGURE 21



UNIT COST VS. FLUE GAS RATE

BASIS: 3% SULFUR COAL - 8600 BTU/KW HR HEAT RATE

NOTE: FLUE GAS RATE IS MEASURED AT INLET CONDITIONS TO SO_2 REMOVAL PLANT and unit costs can then be obtained from Figures 20 and 21. For example, using the conditions cited in the preceding paragraph, wherein an increase in heat rate to 9000 Btu/kwh produces 4.7% more flue gas, the resulting dispersed phase investment, obtained from Figure 20, is about $31.5 million, an increase of about 2.6% over the base case plant. Similarly, the unit operating costs decrease from $4.53/ton to about $4.48/ton, or slightly more than 1% reduction. Thus it appears that modest changes in flue gas volume will have but little effect on SO_2-removal plant investment and operating costs.

3. LOAD FACTOR

A load factor of 60% has been used to calculate the base case unit costs. Load factor is defined as the actual kilowatt hours generated per year divided by the maximum kilowatt hours possible, expressed as a percentage; i.e., load factor = $\frac{\text{Actual kwh/yr}}{\text{Installed kw X 8760 hrs/yr}}$ X 100. Load factor may also be considered as the number of hours per year that the power plant operates at maximum generating capacity; i.e., $\frac{\text{Actual kwh/yr}}{\text{Installed capacity, kw}}$. Thus, in the present study the 60% load factor used to calculate the base case costs is equivalent to 5250 hrs/yr of maximum power generation.

Since the load factor affects the unit operating costs and because it is a variable whose value depends on a number of different parameters, load factors from 50 to 90% have been evaluted for their effect on operating costs. The results of the load factor study are shown in Section C.1, Figures 5, 6, 7 , because the influence of load factor on costs depends largely on the total capital investment even though the variable itself is a function of power plant operation. Since the load factor effect on costs is discussed in Section C.1. it will not be repeated here.

F. PLANT SIZE

1. POWER PLANT SIZE

The base case SO_2-removal plants are sized to treat the maximum flow of flue gases emitted from a 1000 MW coal-fired power plant operating under the specicified conditions; i.e., 13,000 Btu/lb coal, 20% excess air, 8600 Btu/kwh, etc. A 1000 MW size was arbitrarily selected as representative of future installations, but obviously, both larger and smaller plants may be built and it is important to establish the effect of plant size on investment and operating costs. (The possibility of a retrofit SO_2-removal plant is not precluded but since this concept requires a separate study for each specific case it was not considered for the present general evaluation.) Accordingly, two additional sizes have been considered, viz, 500 and 1500 MW, representing a change in the base case size of 1/2 and 1-1/2 respectively.

Investments for both the fluid bed and dispersed phase designs were determined for 500 and 1500 MW plants using base case conditions and Figure presents a plot of change in investment as a function of plant size. (It should be remembered that all investments given in this report are for battery limits only and do not include offsites.) Plant investment for both cases is almost a linear function of plant size over the range, with the fluid bed showing a rate of change slightly higher than the dispersed phase. This near-linearity between plant size and investment is not unexpected since large multiple trains of equipment are involved and only minimal savings are obtained at the larger sizes. The actual change in fluid bed investment from the 1000 MW base is an increase of about $10.5 million for a 1500 MW plant and a $12.3 million decrease for a 500 MW plant. The corresponding values for a dispersed phase plant are $9.2 and $10.2 million. Thus, a 50% increase in plant size raises the investment by about 29-30% while a plant 50% smaller has an investment that is 33-34% less.

FIGURE 22

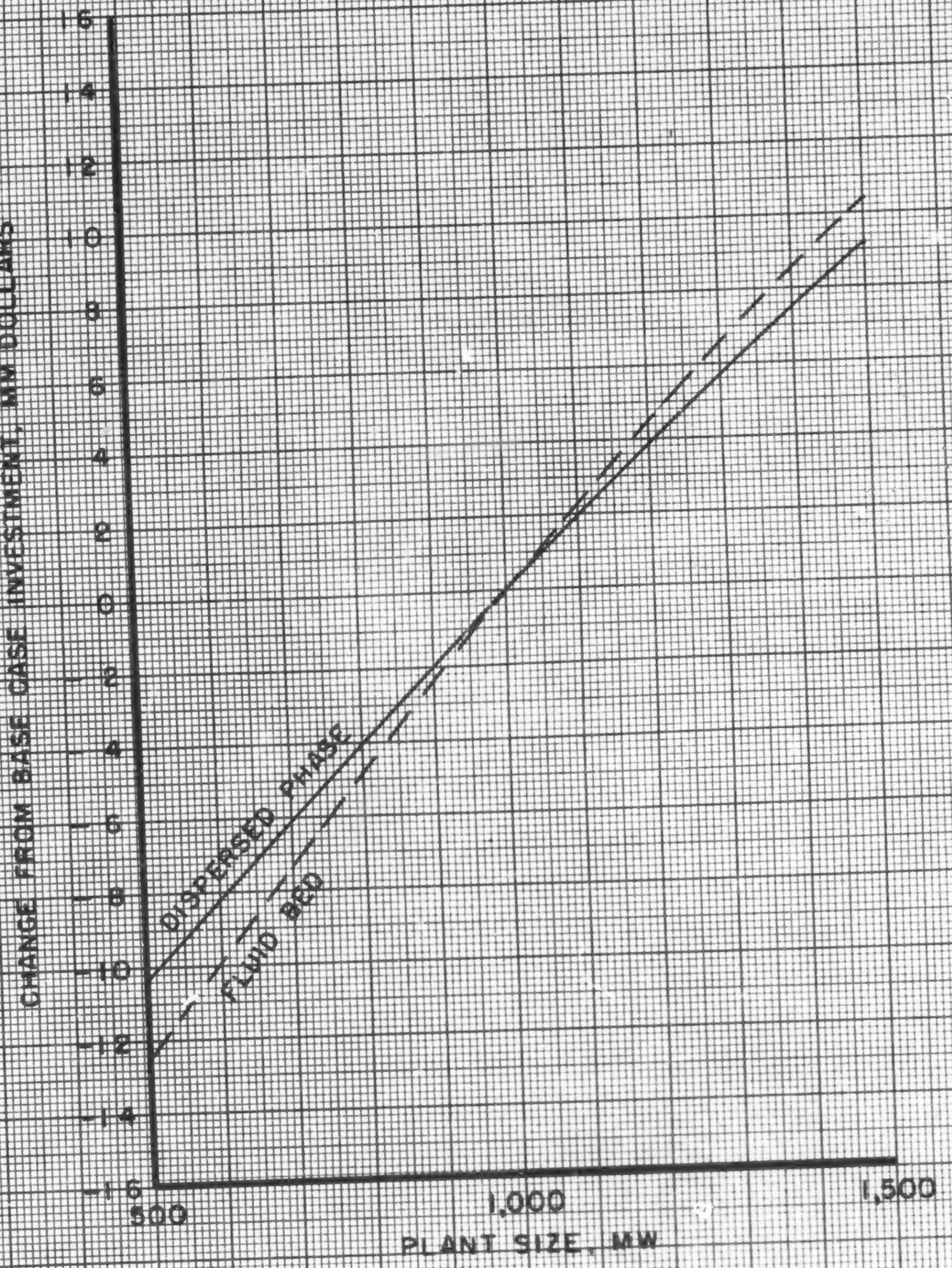
CHANGE IN INVESTMENT FROM BASE CASE VS. PLANT SIZE

3% SULFUR IN COAL

BASE CASE INVESTMENT

DISPERSED PHASE. . . .$30,720,000

FLUID BED. $35,900,000



Unit operating costs of the 500 and 1500 MW plant also have been calculated and the results are shown in Figure 23 as changes from the base case values vs. plant size. It is obvious that operating costs are more sensitive to a decrease in plant size than to an increase; e.g., lowering the plant size from 1000 MW to 500 MW increases the unit cost by $1.40-1.60/ton while a reduction of only $0.55-0.70/ton is obtained when increasing to 1500 MW. The fluid bed costs are slightly more sensitive to plant size than the dispersed phase but even for the largest plant studied (1500 MW), the fluid bed costs are still higher; viz, 4.29 vs. 3.97 dollars per ton of coal. Figure 24 presents the percent change in unit cost from the base case as a function of plant size and it appears that while the lowest unit costs are indicated for a plant size slightly greater than 2000 MW, there is very little decrease in costs as plant size exceeds 1500 MW.

2. SO_2-REMOVAL PLANT SIZE

In evaluating the effect on power plant costs of treating the flue gas to remove SO_2, it has been assumed that the size of the treatment plant should be based on the maximum gas flow rate. Unfortunately, the power plant operates at its maximum generating capacity only about half of the time and, therefore, the SO_2-removal plant also has a low operating efficiency. If, however, size of the flue gas treatment plant were based on the gas volume produced during periods of off-peak power generation, then the equipment required would be smaller and the operating efficiency would increase. Since this latter scheme appeared to offer potential for substantial cost savings over the base case it was decided to evaluate the economic aspects of a half-size SO_2-removal plant.

With a half-size gas treatment plant, only half of the flue gas can be treated during periods of maximum power generation

FIGURE 23

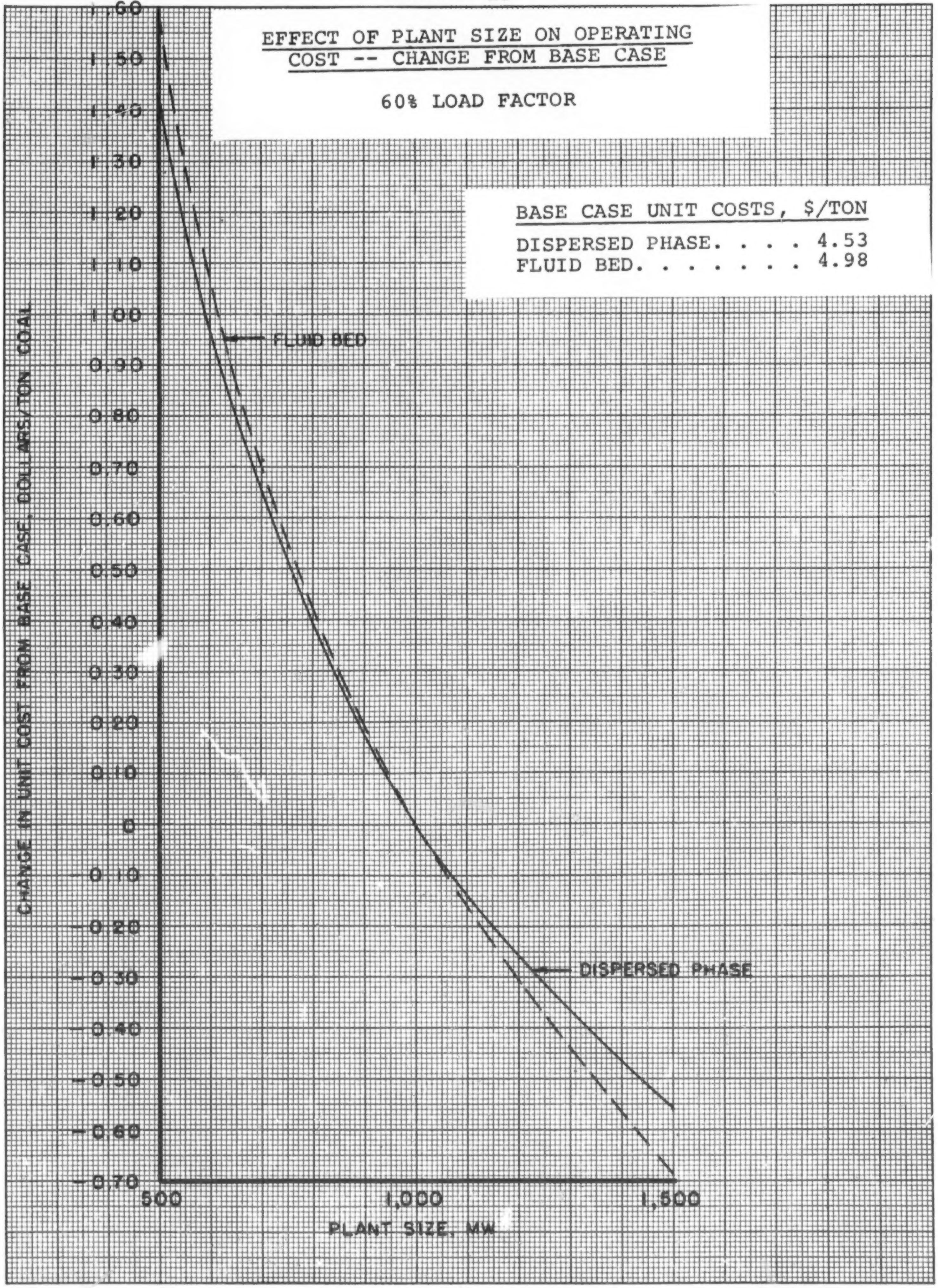
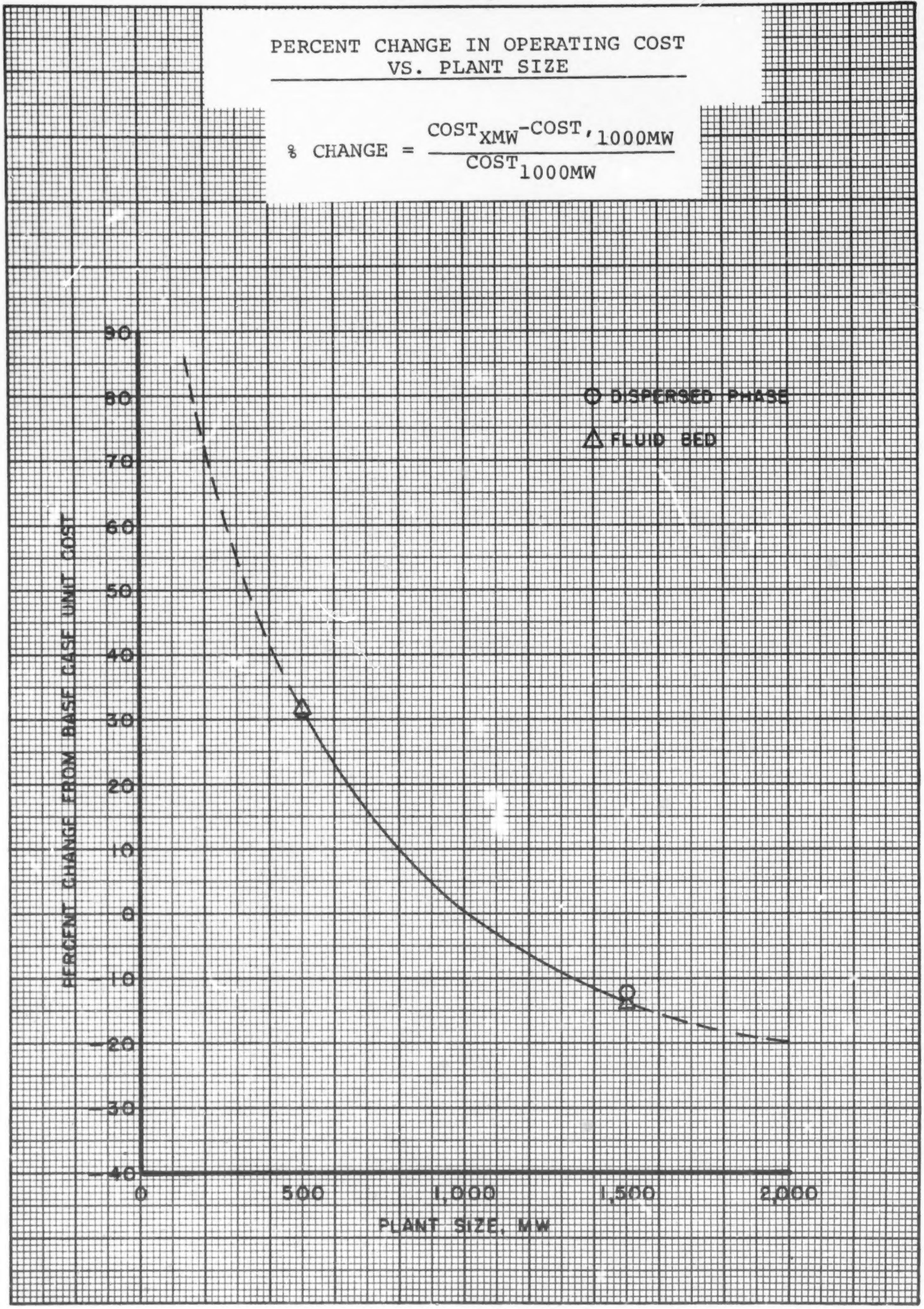

EFFECT OF PLANT SIZE ON OPERATING COST -- CHANGE FROM BASE CASE
60% LOAD FACTOR
BASE CASE UNIT COSTS, $/TON
DISPERSED PHASE. . . . 4.53
FLUID BED. 4.98
FLUID BED
DISPERSED PHASE
CHANGE IN UNIT COST FROM BASE CASE, DOLLARS/TON COAL
1.60
1.50
1.40
1.30
1.20
1.10
1.00
0.90
0.80
0.70
0.60
0.50
0.40
0.30
0.20
0.10
0
-0.10
-0.20
-0.30
-0.40
-0.50
-0.60
-0.70
500
1,000
1,500
PLANT SIZE, MW

FIGURE 24



PERCENT CHANGE IN OPERATING COST VS. PLANT SIZE

$$\% \text{ CHANGE} = \frac{\text{COST}_{\text{XMW}}-\text{COST},_{\text{1000MW}}}{\text{COST}_{\text{1000MW}}}$$

but it is assumed that air quality standards can be maintained by burning low sulfur coal in one half of the power plant. High sulfur coal can be used in the other half of the power plant since the SO_2 is being removed from the flue gases. During periods of off-peak power generation only high sulfur coal is burned because all of the flue gas produced under these conditions can be treated. If this mode of operation is feasible, then low sulfur coal will be required for about 25% of the time while high sulfur coal can be burned the other 75%. In the following economic evaluations no allowances have been made for any cost penalties owing to the part time use of low sulfur coal; i.e., no extra investment was included for storage and handling facilities needed to alternately burn both low and high sulfur coal in the same unit, nor was any allowance made for the probable higher price of low sulfur coal.

The investments and operating costs for three different half-size SO_2-removal plants have been prepared, the bases being:

Case 1:	600°F sorption/1400°F regeneration, dispersed phase design
Case 2:	300°F sorption/1400°F regeneration dispersed phase design
Case 3:	300°F sorption/1400°F regeneration fluid bed design

An evaluation for the fixed bed design was not made mainly because it does not offer the same potential for cost reduction as do the other two cases. That is, a half-size plant for either the dispersed phase or fluid bed permits the deletion of one entire train of equipment whereas a half-size fixed bed plant requires the same, albeit smaller sized, equipment.

Schematic block-type layouts of the half size SO_2-removal plants are shown in Figures 25 and 26. Figure 25 shows two possible arrangements for the Case 1 conditions of 600°F sorption/ 1400°F regeneration; i.e., the SO_2-removal plant is located

3 OF 4

PB
200 882

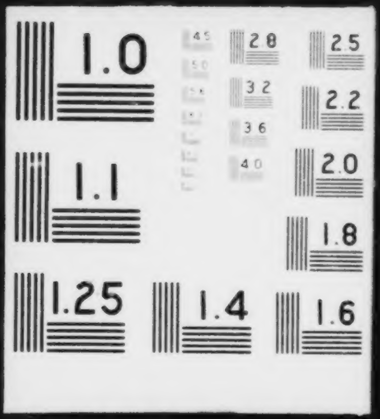
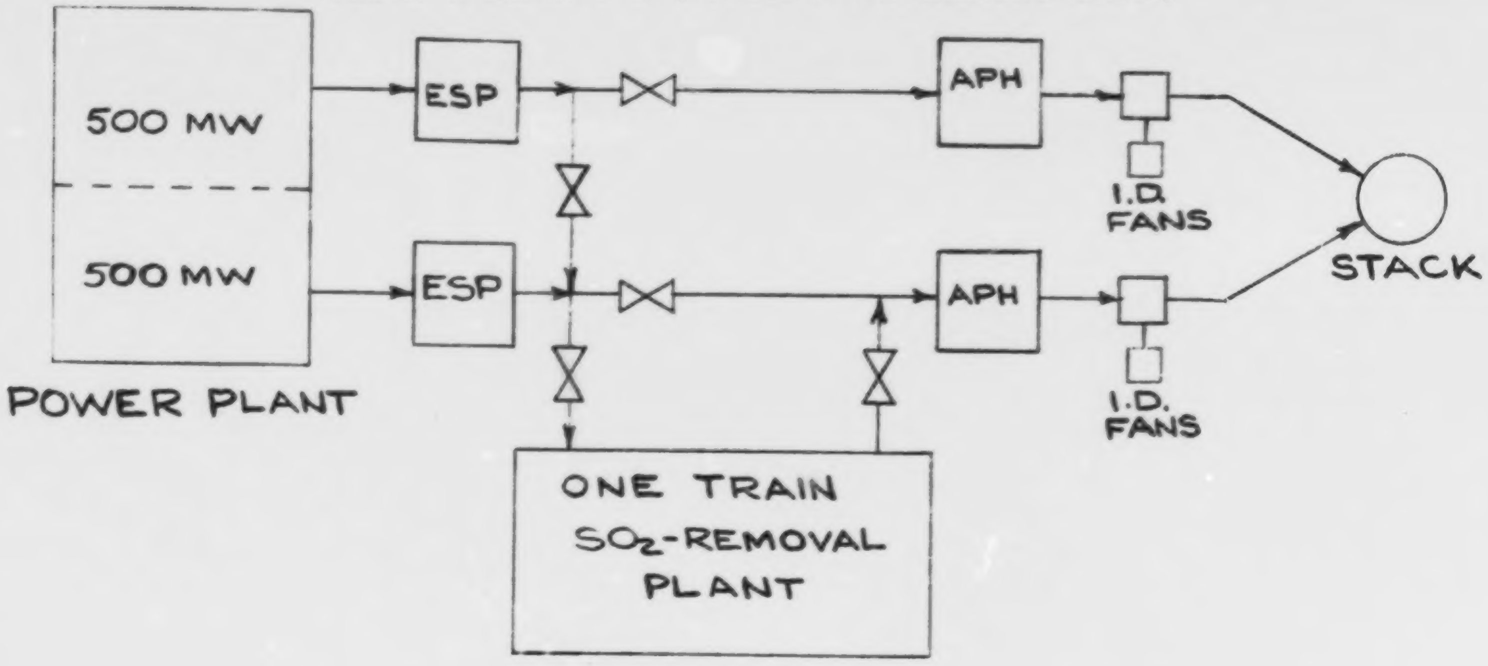

1.0
4.5
2.8
2.5
5.0
3.2
2.2
3.6
4.0
2.0
1.1
1.8
1.25
1.4
1.6

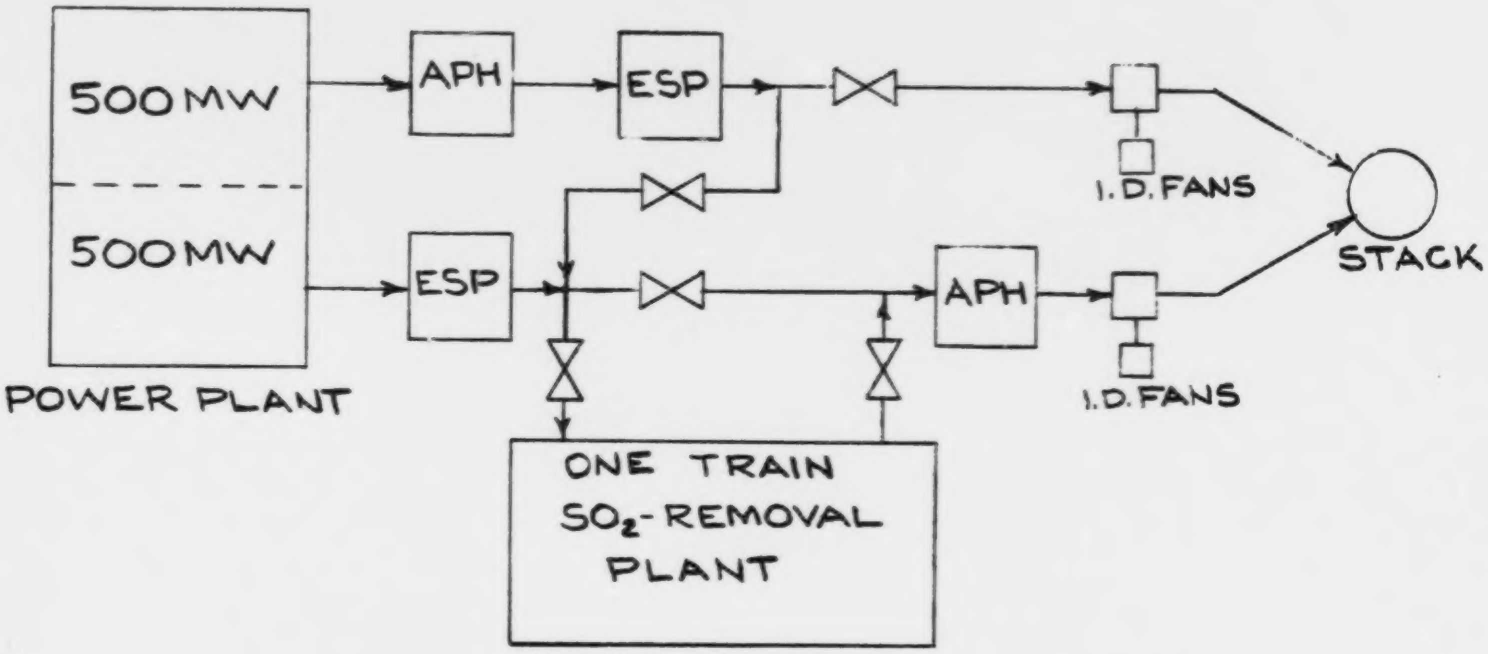
FIGURE 25

SCHEMATIC LAYOUT OF HALF-SIZE SO_2-REMOVAL PLANT
(600°F SORPTION/1400°F REGENERATION)


500 MW
500 MW
POWER PLANT
ESP
ESP
APH
APH
I.D. FANS
I.D. FANS
STACK
ONE TRAIN
SO_2-REMOVAL
PLANT

CASE 1

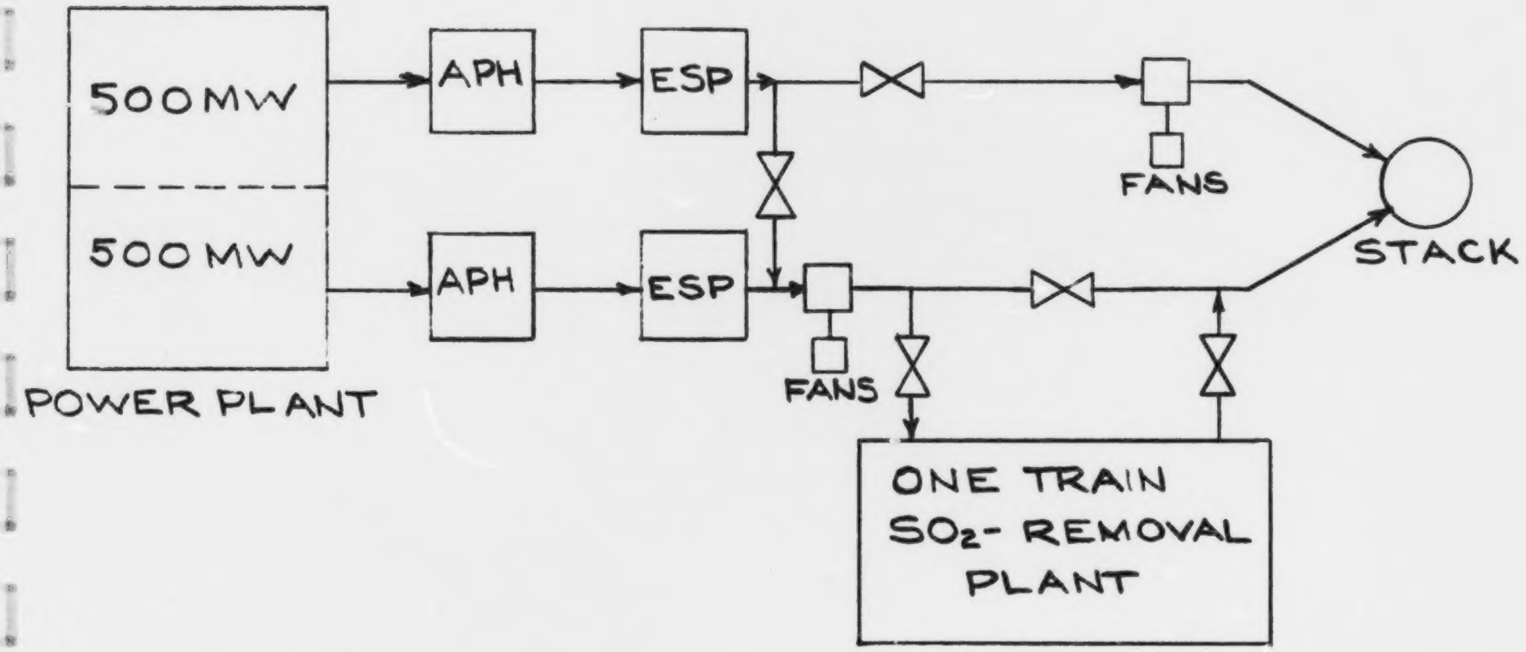
NOTE: ESP - ELECTROSTATIC PRECIPITATOR
APH - AIR PREHEATER


500MW
500MW
POWER PLANT
APH
ESP
ESP
APH
I.D. FANS
I.D. FANS
STACK
ONE TRAIN
SO_2-REMOVAL
PLANT

CASE 1-A

FIGURE 26

SCHEMATIC LAYOUT OF HALF-SIZE SO_2-REMOVAL PLANT
(300°F SORPTION/1400°F REGENERATION)



500MW
500 MW
POWER PLANT
APH
APH
ESP
ESP
FANS
FANS
STACK
ONE TRAIN
SO2- REMOVAL
PLANT

NOTE: ESP- ELECTROSTATIC PRECIPITATOR
APH - AIR PREHEATER

CASES 2 & 3

upstream of the air preheater. The valving arrangement shown in Figures 25 and 26 allows the plant to be operated in three ways:

1. Applies to periods when the power plant is operating at 50% or less of capacity. During this time all flue gas is routed through the SO_2-removal plant.

2. For periods when the power plant is operating at greater than 50% of capacity. During this time at least half of the flue gas is desulfurized in the SO_2-removal plant while the remainder is routed directly to the stack.

3. The isolation valves permit the SO_2-removal plant to be blocked out of operation for maintenance purposes while all flue gases are routed directly to the stack.

The economics of the Case 1 (600°/1400°) half-size dispersed phase SO_2-removal plant are shown in Table 22 based on the Case 1 layout shown in Figure 25. The dispersed phase plant was chosen for economic evaluation rather than the fixed or fluid bed designs because it seemed to have the best chance for the lowest unit cost. Total fixed capital investment for the half-size dispersed phase plant is estimated to be $22,200,000, a savings of $8,520,000 (28%) over the full size base case plant. This figure includes the cost for the five valves shown in Figure 25 and also the incremental cost of the electrostatic precipitators and air preheaters. Total annual operating costs for this plant amounts to $5,757,100/yr, which is equivalent to unit costs of 3.32/ton coal (12.77¢/MM Btu or 1.10 mills/kwh). These unit costs represent a 27% reduction from the base case cost of $4.53/ton coal for a full size dispersed phase SO_2-removal plant.

An alternative design possibility is shown in Figure 25, Case 1-A. In this design one electrostatic precipitator has

TABLE 22

DISPERSED PHASE - PROCESS ECONOMICS
HALF-SIZE SO_2 REMOVAL PLANT
600°F SORPTION/1400°F REGENERATION

	DIRECT COST	$/YR
1.	OPERATING LABOR (3 MEN/SHIFT @$3.50/HR)	92,000
2.	SUPERVISION	15,000
3.	MAINTENANCE - 5% OF FCI*	1,110,000
4.	SUPPLIES - 15% OF MAINTENANCE	166,500
5.	UTILITIES: a) COOLING WATER @ 5¢/M GAL	59,300
	b) NATURAL GAS @ 27¢/MM BTU	730,000
	c) POWER @ 0.6¢/KWH	289,000
6.	HEAT (CREDIT) @ 25¢/MM BTU	(219,800)
7.	STEAM (CREDIT) @ 25¢/M LBS	(76,100)
8.	TOTAL DIRECT COST	2,165,900
	INDIRECT COST	
9.	PAYROLL OVERHEAD - 20% of (1+2)	21,400
10.	PLANT OVERHEAD - 50% OF (1+2+3+4)	691,800
11.	TOTAL INDIRECT COST	713,200
	FIXED COST	
12.	CAPITAL CHARGES @ 13% OF FCI* (DEPRECIATION, INSURANCE, TAXES, COST OF CAPITAL, ETC.)	2,886,000
13.	SUBTOTAL OPERATING COST (8+11+12)	5,765,100
14.	ATTRITION LOSS 2,780,000 LBS/YR @ 25¢/LB	695,000
15.	SULFUR CREDIT 35,150 TONS/YR @ $20/TON	(703,000)
16.	TOTAL (13+14+15)	5,757,100**
17.	UNIT COST: a) $/TON COAL (1,736,537 TONS/YR)	3.32
	b) ¢/MM BTU (26MM BTU/TON)	12.77
	c) MILLS/KWH	1.10

*FCI (FIXED CAPITAL INVESTMENT) = $22,200,000

**DOES NOT INCLUDE THE COST OF LOW SULFUR COAL.

NOTE: THE SO_2 PLANT OPERATES AT 90% LOAD FACTOR OR 7875 HOURS PER YEAR
THE POWER PLANT OPERATES AT 60% LOAD FACTOR OR 5250 HOURS PER YEAR.

been relocated so that its incremental cost is reduced from \$3,656,000 (Case 1) to \$2,119,000 (Case 1-A). One disadvantage to the Case 1-A design is that the SO_2-removal plant will sometimes receive 600°F flue gas and sometimes mixed gases at some lower temperature. During the lower temperature operation, the natural gas requirements of the SO_2-removal plant will increase, thus at least partially offsetting the savings obtained from the lower investment. Therefore, while detailed economics have not been worked out to determine the unit costs for the Case 1-A design, no large savings are expected.

Since substantial savings were obtained for the full size plant when the sorption temperature was lowered from 600°F to 300°F, it was decided to evaluate a half-size plant for these conditions. The economics for the 300°F sorption designs (Cases 2 and 3, Figure 26) are shown in Tables 23 and 24.

Referring to Table 23(i.e., the 300°/1400° dispersed phase case), plant investment is \$16,450,000, which is a reduction of \$14,270,000 (46%) over the corresponding full size plant, and the operating costs are reduced to \$2.68/ton coal (-41%). As noted on Table 23 the operating costs are based on a power plant load factor of 60% while it has been assumed that the SO_2-removal plant will operate 90% of the time.

Table 24 shows the economics for the half-size, 300°/1400° fluid bed design. For this alternative the investment is reduced to \$17,840,000, which is \$18,060,000 less (>50%) than the base case, while operating costs decrease to \$2.64/ton coal (-47%). The same load factors are used as before, viz, 60% and 90% for the power plant and SO_2-removal plant, respectively.

The preceding costs shown for the half-size plants possibly could be reduced further if a 600°F regeneration temperature were used. Based on the results obtained for the full size plants, it appears that lowering the regeneration from 1400°F

TABLE 23

DISPERSED PHASE - PROCESS ECONOMICS
HALF-SIZE SO_2-REMOVAL PLANT
300° SORPTION/1400° REGENERATION

	DIRECT COST	$/YR
1.	OPERATING LABOR (3 MEN/SHIFT @$3.50/HR)	92,000
2.	SUPERVISION - 15% OF LABOR	13,800
3.	MAINTENANCE - 5% OF FCI*	822,500
4.	SUPPLIES - 15% OF (3)	123,400
5.	UTILITIES: a. COOLING WATER @5¢/M GAL	59,300
	b. NATURAL GAS @27¢/MM BTU	734,000
	c. POWER @ 0.6¢/KWH	296,100
6.	HEAT (CREDIT) @ 25¢/MM BTU	(87,800)
7.	STEAM (CREDIT) @25¢/M LBS	(76,200)
8.	TOTAL DIRECT COST	1,977,100
	INDIRECT COST	
9.	PAYROLL OVERHEAD 20% OF (1+2)	21,200
10.	PLANT OVERHEAD 50% of (1+2+3+4)	525,900
11.	TOTAL INDIRECT COST	547,100
	FIXED COST	
12.	CAPITAL CHARGES @ 13% (DEPRECIATION, INSURANCE, TAXES, COST OF CAPITAL, ETC.)	2,138,500
13.	SUB TOTAL OPERATING COSTS (8+11+12)	4,662,700
14.	ATTRITION LOSS (2,780,000 LB/YR)	695,000
15.	SULFUR CREDIT (35,150 TONS/YR)	(703,000)
16.	TOTAL COST (13+14+15)	4,654,700 **
17.	UNIT COSTS: a. $/TON COAL (1,736,537 TONS/YR)	2.68
	b. ¢/MM BTU (26 MM BTU/TON)	10.31
	c. MILLS/KWH	0.89

* FCI (FIXED CAPITAL INVESTMENT) = $16,450,000

** DOES NOT INCLUDE THE COST OF LOW SULFUR COAL

NOTE: THE SO_2 PLANT OPERATES AT 90% LOAD FACTOR OR 7875 HOURS/YR
THE POWER PLANT OPERATES AT 60% LOAD FACTOR OR 5250 HOURS/YR

TABLE 24

FLUID BED - PROCESS ECONOMICS
HALF-SIZE SO_2-REMOVAL PLANT
300°F SORPTION/1400°F REGENERATION

		$/YR
DIRECT COST		
1.	OPERATING LABOR (4 MEN/SHIFT @$3.50/HR)	123,000
2.	SUPERVISION - 15% OF LABOR	18,500
3.	MAINTENANCE - 5% OF FCI*	892,000
4.	SUPPLIES - 15% OF (3)	133,800
5.	UTILITIES: a. COOLING WATER @5¢/1000 GAL	59,300
	b. NATURAL GAS @ 27¢/MM BTU	673,700
	c. POWER @0.6¢/KWH	412,900
6.	HEAT (CREDIT) @25¢/MM BTU	(59,500)
7.	STEAM (CREDIT) @25¢/M LBS	(76,100)
8.	TOTAL DIRECT COST	2,177,600
INDIRECT COST		
9.	PAYROLL OVERHEAD - 20% OF (1+2)	28,300
10.	PLANT OVERHEAD - 50% OF (1+2+3+4)	583,700
11.	TOTAL INDIRECT COST	612,000
FIXED COST		
12.	CAPITAL CHARGES @13% (DEPRECIATION, INSURANCE, TAXES, COST OF CAPITAL, ETC.)	2,319,200
13.	SUBTOTAL OPERATING COSTS	5,108,800
14.	ATTRITION LOSS (748,000#/YR)	187,000
15.	SULFUR CREDIT (35,150 TON/YR)	(703,000)
16.	TOTAL	4,592,800**
17.	UNIT COSTS a. $/TON COAL (1,736,537/TONS YR)	2.64
	b. ¢/MM BTU (26 MM BTU/TON)	10.17
	c. MILLS/KWH	0.87

* FCI (FIXED CAPITAL INVESTMENT) = $17,840,000

** DOES NOT INCLUDE THE COST OF LOW SULFUR COAL

NOTE: THE SO_2 PLANT OPERATES AT 90% LOAD FACTOR OR 7875 HOURS/YR
THE POWER PLANT OPERATES AT 60% LOAD FACTOR OR 5250 HOURS/YR

to 600°F produces savings of 2-10%. However, since no 600°F regeneration data are available, it was felt that evaluating the effect of low regeneration temperature on the half-size plant costs was not justified, particularly in view of the potentially modest savings.

It is interesting to note that while the two 300°F/1400°F half-size plants discussed above show the lowest costs of all the cases studied, the by-product sulfur credit is still much less than the SO_2-removal plant costs. A sulfur break even selling price (see Section D.3.) was calculated for these half-size plants and for the dispersed phase is \$152/ton while the fluid bed value is \$151/ton. The corresponding values for the full size plants are \$158/ton and \$159/ton. Thus, despite the large savings realized in unit costs, the sulfur break even price is reduced only slightly.

G. SORBENT

1. SORBENT LOADING

The effects of sorbent loading (i.e., weight percent sulfur on the sorbent) on process economics have been determined for the dispersed phase and fluid bed designs. For the dispersed phase, sorbent loadings of 1.5 and 6.0% S have been compared to the base case loading of 3.0% S, regenerating in each instance to the 0.5% S level of the base case. For the fluid bed, sorbent loadings of 5.0% and 9.0% S were compared to the base case loading of 7.0% S, regenerating to 2.6% S in each case.

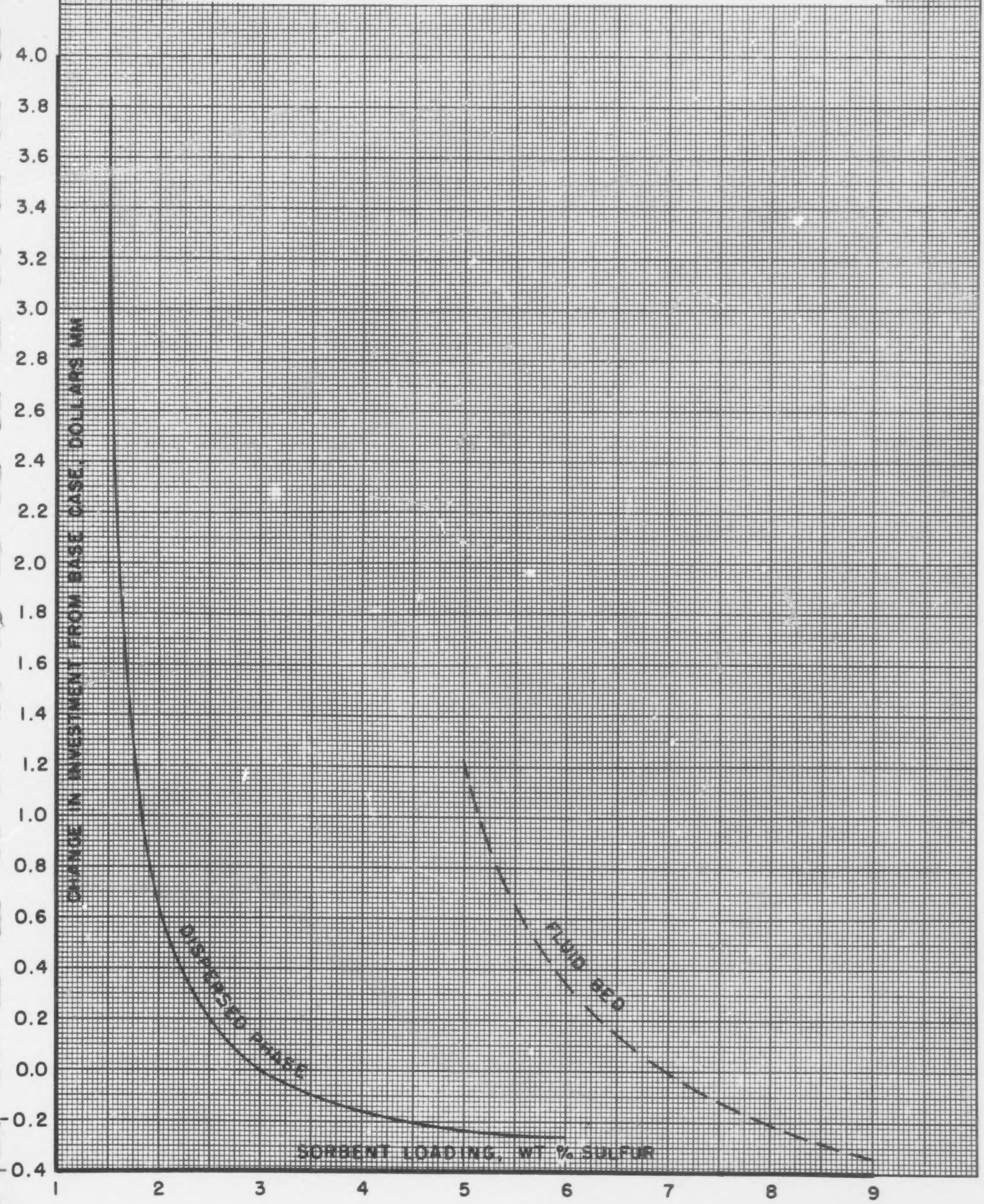
Figure 27 shows the effect of sorbent loading on plant investment. It is obvious that the dispersed phase design is far more sensitive to changes in sorbent loading than the fluid bed design. The changes in investment for the low and high sorbent loadings are +$3,830,000 (+12.5%) and -$260,000 (-0.8%) respectively, for the dispersed phase, while comparable figures for the fluid bed design are +$1,220,000 (+3.4%) and -$340,000 (-0.9%). It is also apparent that both designs are more sensitive to decreases in sorbent loading than to increases. This means that whereas decreases in sorbent loadings require substantial increases in investment, only small savings are achieved in going to higher loadings.

For the dispersed phase, the major effects resulting from lowering the sorbent loading to 1.5% are increases in the solids circulation rate and the solids reheat duty. As a result, most of the equipment in the sorption/ regeneration unit increases in size, with the exception of the reactor, which becomes smaller because of the reduced reaction time required to reach a 1.5% loading. The overall net effect caused by these changes is an increase in the investment for the sorption/regeneration unit of about

FIGURE 27

EFFECT OF SORBENT LOADING ON TOTAL INVESTMENT

1000 MW POWER PLANT



$1,600,000. Owing to larger vessels sizes, the sorbent inventory also increases, resulting in a further increase in investment of about $700,000.

An increase of about $500,000 in investment is required for the power plant electrostatic precipitator owing to the larger amounts of combustion and transport air used. This increased air flow also results in an incremental investment of about $1,100,000 for the air preheater.

No change in investment is required for the regeneration gas plant or the Claus plant and its auxiliaries. Changing the sorbent loading does not affect the amount of SO_2 removed from the flue gas, which remains the same as all cases. Hence, the regeneration gas requirements and the Claus plant feed gas remain constant.

Increasing the sorbent loading to 6.0% produces, in general, opposite effects to those discussed above. The solids circulation rate and solids reheat duty decrease, resulting in a decrease in size of most of the process equipment. However, a substantial increase in reactor height is required because of the longer reaction time required to achieve the 6.0% loading. This additional reactor height not only increases the reactor investment, but, because of the resulting increase in the quantity of large-diameter flue gas ducting, it also substantially increases the investment of the ductwork. This latter increase more than compensates for the smaller size of the rest of the process unit with the result that the process unit investment actually increases by about $400,000. This does not include any increase in investment owing to increased pressure drop of the flue gas through the taller reactor; i.e., the flue gas blower discharge pressure has been kept constant. Because of the lower solids circulation rate, hence less holdup in the heater and regenerator, there is a modest reduction of about $100,000 in sorbent inventory. The cost of the power plant electrostatic precipitator is also reduced, by about $600,000, since the flue gas quantity is somewhat lower than in the base case. The net result of all of the above investment changes is a reduction in plant investment of $260,000.

For the fluid bed case, similar effects are produced. Lowering the sorbent loading to 5.0% S increases the solids circulation rate and reheat duty, but decreases the required residence time of the solids in both the reactor and regenerator. As a result of the increased solids circulation rate, both the solids heater and the air compressor increase significantly in cost. It might be expected that this would be at least partialy offset by a decrease in costs for both the reactor and regenerator, since the theoretical residence time requirements are reduced for both vessels. This is not the case however. In the reactor, the minimum bed height needed for operability is greater than that required by kinetics and and consequently there is no change in reactor size. The regenerator does decrease somewhat in height, but since all internals, nozzles, cyclones, etc remain constant, the cost reduction is small.

The only other significant cost change is that of the air preheater. This increases by $200,000 due to the increase in process air requirements.

Increasing the sorbent loading to 9% produces a modest decrease in plant investment. In this case, the solids circulation rate and reheat duty decrease, while the required residence time of the solids in both the reactor and regenerator increase. These changes are relatively small, however, and result in small reductions in equipment costs, the largest (about $175,000) being that of the air compressor.

The effect of sorbent loading on operating costs are shown in Figures 28 and 29 for the dispersed phase and fluid bed designs respectively. Again, the dispersed phase is more sensitive than

FIGURE 28

EFFECT OF SORBENT LOADING ON OPERATING COSTS - DISPERSED PHASE

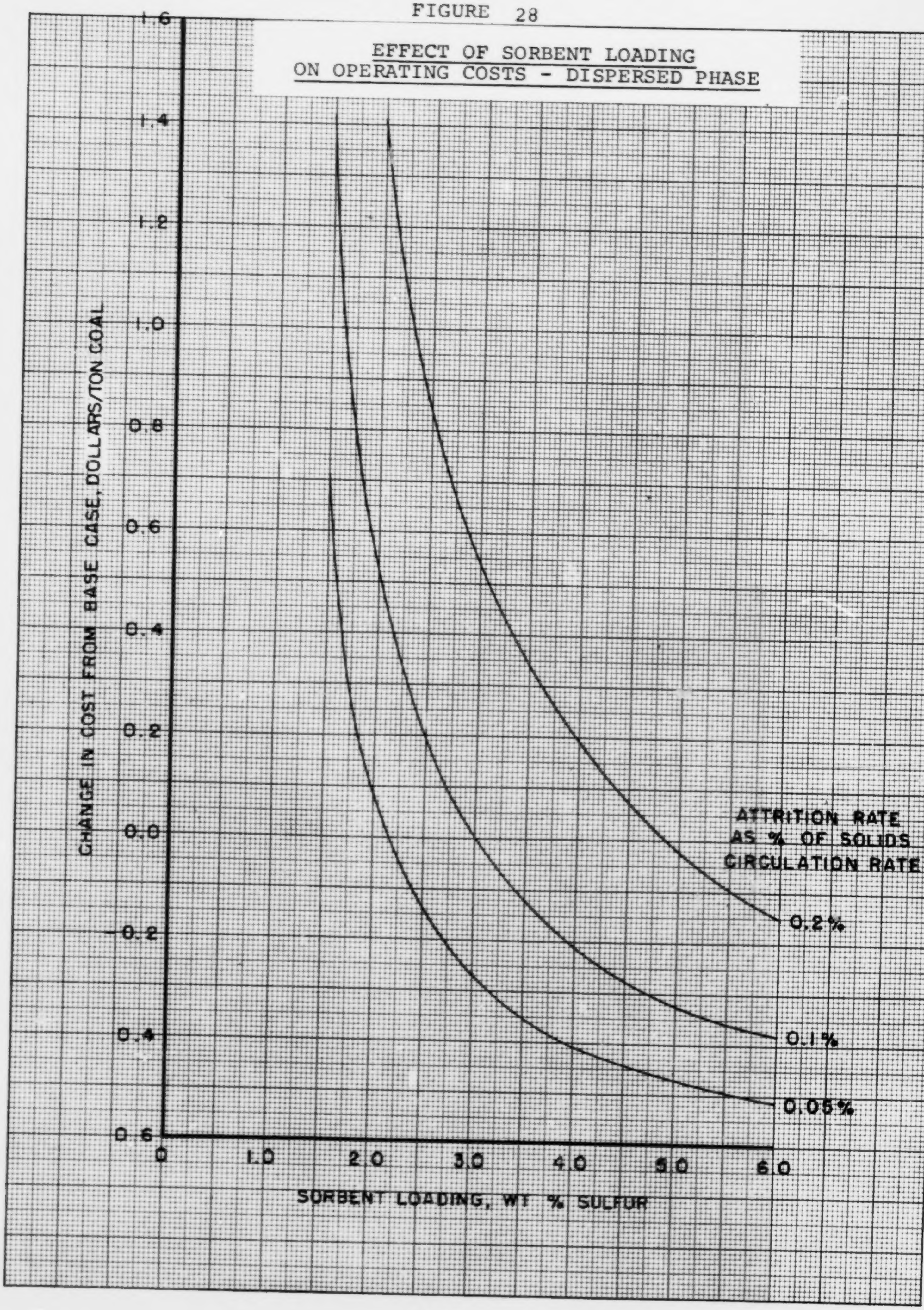
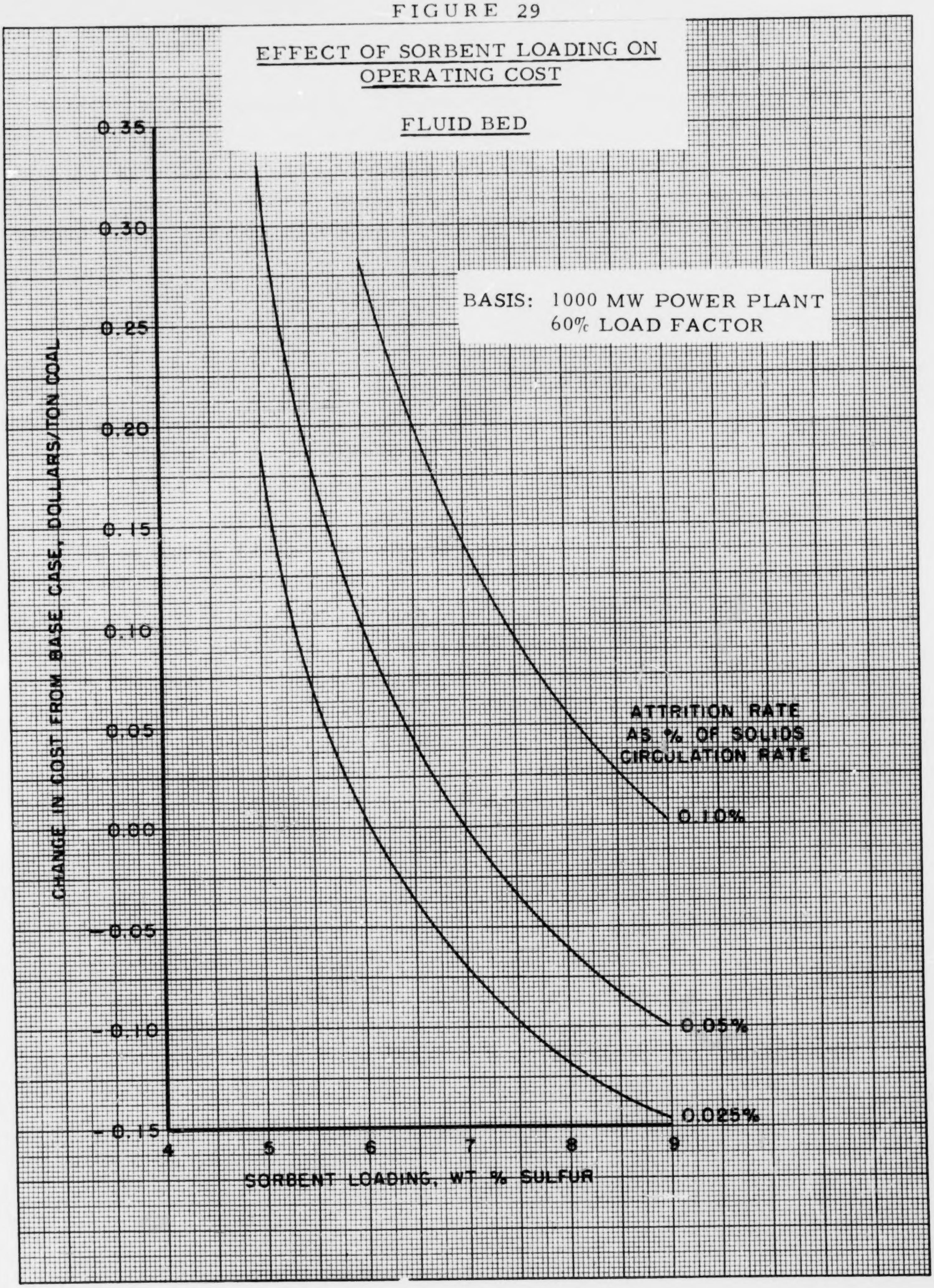


FIGURE 29


EFFECT OF SORBENT LOADING ON OPERATING COST
FLUID BED
BASIS: 1000 MW POWER PLANT
60% LOAD FACTOR
CHANGE IN COST FROM BASE CASE, DOLLARS/TON COAL
0.35
0.30
0.25
0.20
0.15
0.10
0.05
0.00
-0.05
-0.10
-0.15
4
5
6
7
8
9
SORBENT LOADING, WT % SULFUR
ATTRITION RATE AS % OF SOLIDS CIRCULATION RATE
0.10%
0.05%
0.025%

the fluid bed to changes in sorbent loading, and both designs are more sensitive to decreases in loadings than to increases. At the low and high sorbent loadings respectively, the changes in unit cost are +$1.42/ton coal and -$0.38/ton coal for the dispersed phase and +$0.33/ton coal and -$0.10/ton coal for the fluid bed.

Although various costs, such as natural gas charges, power costs, heat credits, capital charges, etc. are affected, the most important cost change is the attrition loss. This loss was calculated on the basis of 0.1% of the solids throughput rate for the dispersed phase and 0.05% for the fluid bed.

The effect of two different attrition rates on plant costs is also shown in Figures 28 and 29, viz, one-half and twice the base case solids throughput rate. Because of the major economic effect of attrition rate, the variation of plant cost with attrition loss shown for the base case cannot be used if the solids throughput rate is appreciably different from that of the base case. If the effect of attrition rates other than those shown is desired, the values from Figures 28 and 29 may be easily extrapolated since the relationship between attrition loss and plant cost is a straight line function at any given sorbent loading.

The apparent mechanism by which attrition occurs (i.e., a chemical weakening of the sorbent structure coupled with the stresses imposed by fluidizing and transporting the solids) indicates that an increase in attrition rate of 0.05% or more is quite possible. Thus, the operating costs are much more sensitive to attrition rate than sorbent loading. It is apparent from the curves that as the sorbent loading decreases, hence solids circulation rate increases, the effect of attrition loss on unit cost is more pronounced.

2. ATTRITION RATE/SORBENT PRICE

The effects of attrition rate and sorbent cost on plant operating costs have been determined for the three base case designs. Figures 30, 31, and 32 are plots of the attrition rate vs. annual operating cost, in dollars per ton of coal, with parameters of sorbent cost and sulfur credit, for the dispersed phase, fluid bed, and fixed bed cases respectively. Attrition rate is expressed as a percentage of solids throughput for the dispersed phase and fluid bed cases whereas years of sorbent life are used for the fixed bed system.

Examination of Figures 30-32 shows that attrition rate and sorbent price have the most pronounced effect on the dispersed case economics and the least effect on the fixed bed case. That is, as the attrition rate is increased, the sorbent cost has a greater effect on the dispersed phase operating cost than it does on the other two systems. In addition, the rate of attrition has a much greater effect on dispersed phase economics than on the other two systems, particularly at sorbent prices of about 25 cents per pound and higher. In fact, as can be seen from Figure 32, fixed bed operating costs are essentially independent of attrition loss beyond a sorbent life of about three years (equivalent to a dispersed phase attrition rate of 0.1% in the base case economics).

Inspection of Figures 30 and 31, at a 0.0% attrition rate, and Figure 32, at sorbent life of 8 years (approximating a 0.0% attrition loss for the fixed bed case) shows that the curves for various sorbent costs do not originate at a common point. This is so because the cost of the sorbent inventory is included in the total plant investment. Capital charges on this inventory account for the differences in annual operating costs at 0.0% attrition rate.

FIGURE 30

EFFECT OF ATTRITION RATE ON DISPERSED PHASE OPERATING COSTS

NOTE: - 1000 MW POWER PLANT
- CAPITAL CHARGE RATE IS 13% FIXED CAPITAL INVESTMENT/YR
- 60% LOAD FACTOR

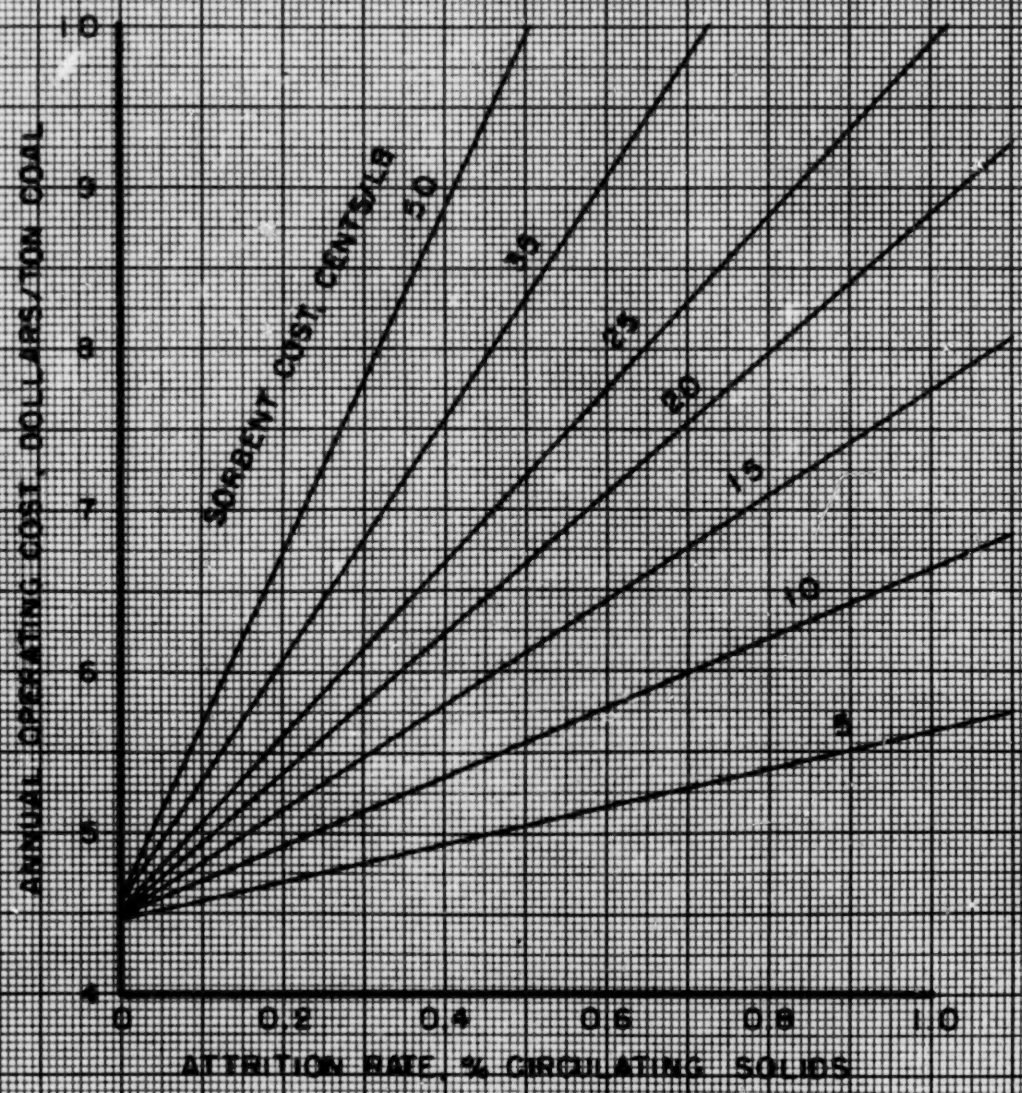
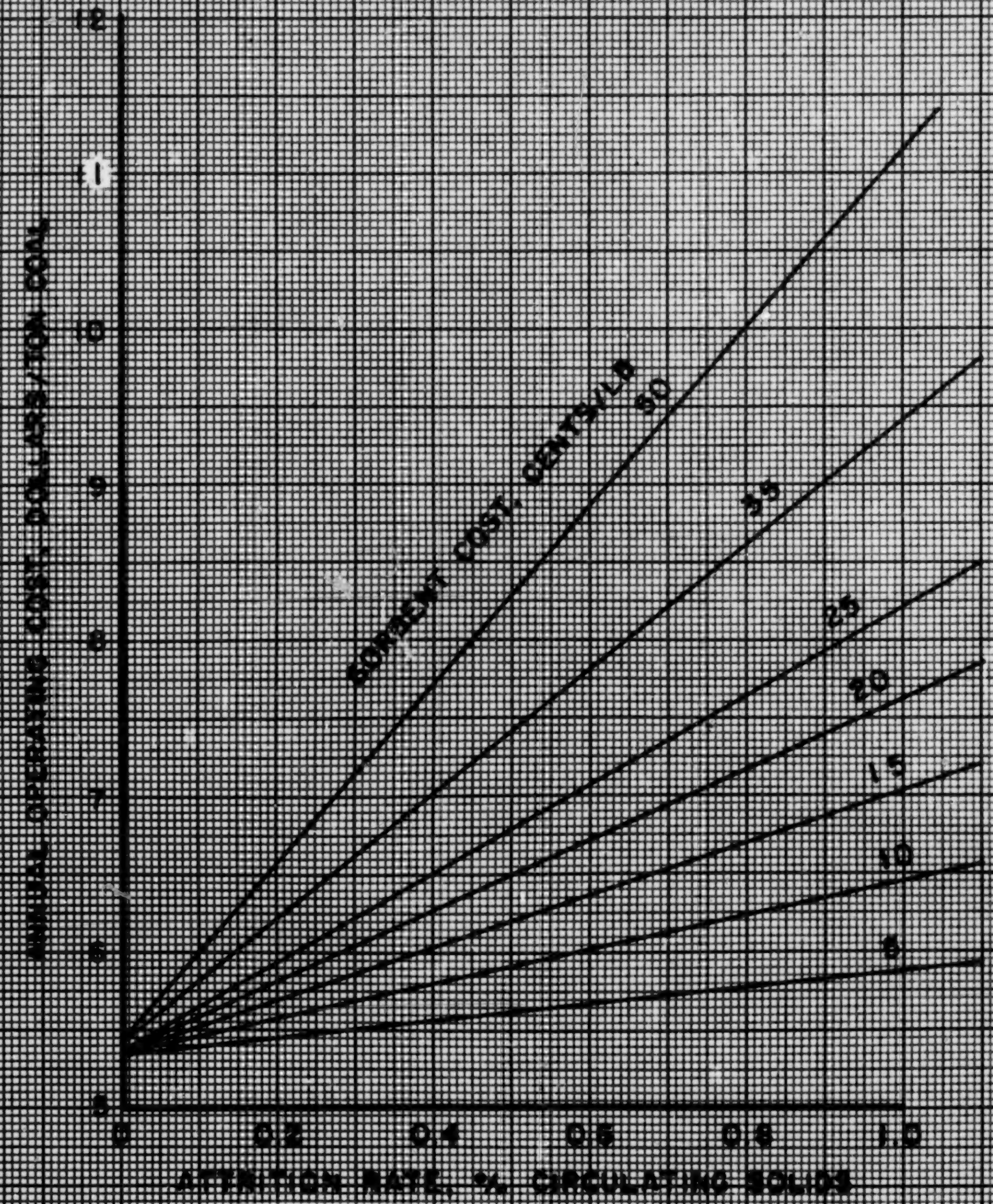


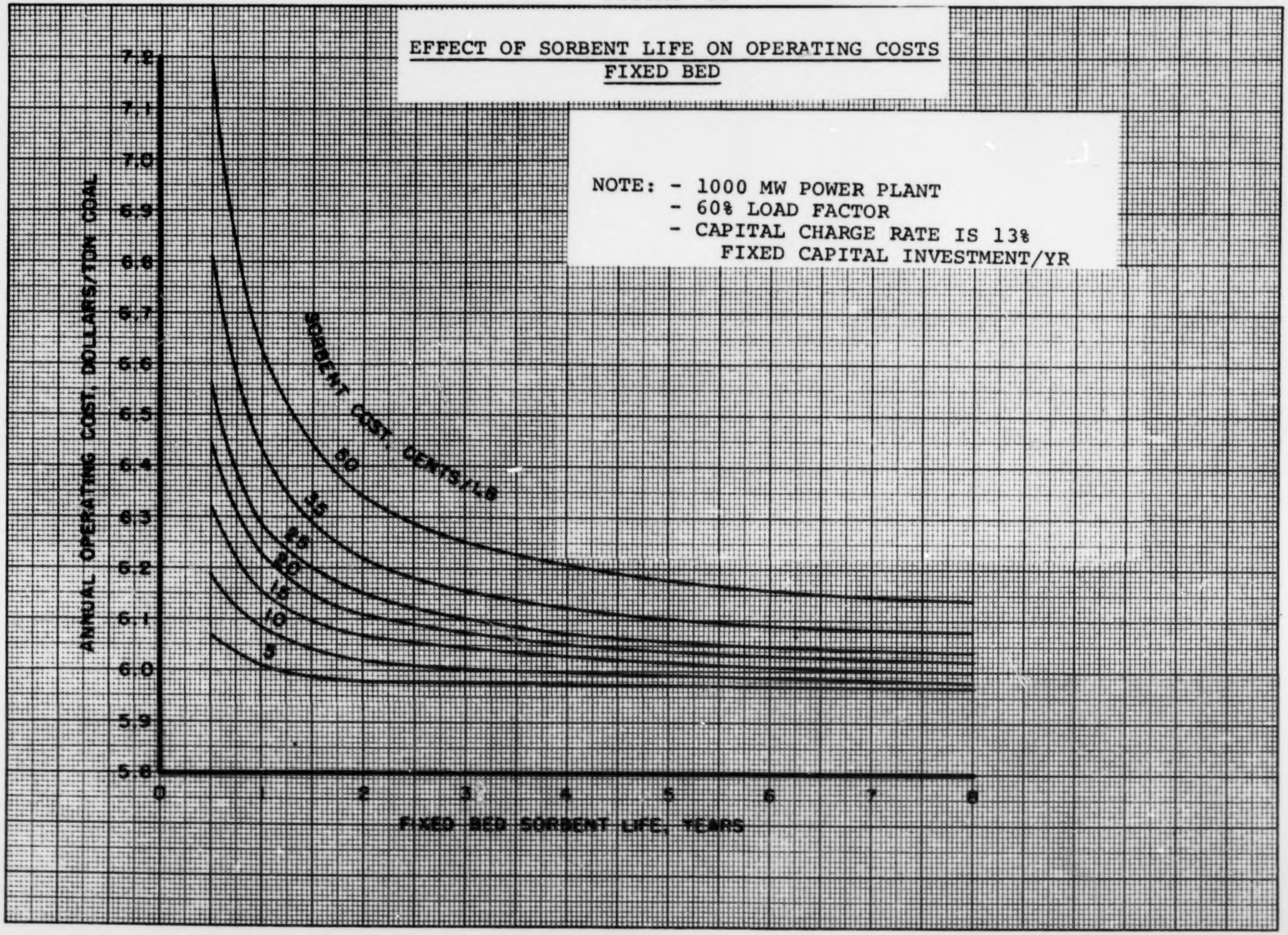
FIGURE 31

EFFECT OF ATTRITION RATE ON FLUID BED OPERATING COSTS



BASIS: - 1000 MW POWER PLANT
- CAPITAL CHARGE RATE IS 13% FIXED CAPITAL INVESTMENT/YR
- 60% LOAD FACTOR

FIGURE 32



EFFECT OF SORBENT LIFE ON OPERATING COSTS
FIXED BED

NOTE: - 1000 MW POWER PLANT
- 60% LOAD FACTOR
- CAPITAL CHARGE RATE IS 13% FIXED CAPITAL INVESTMENT/YR

H. REGENERATION GAS PRODUCTION

1. REFORMER DESIGN

The reducing gas used to regenerate sulfur loaded sorbent from the reactor is a hydrogen- and carbon monoxide-rich gas produced by steam-reforming natural gas. The choice of this process was made following a brief consideration of other alternatives (i.e., produced gas, water gas, etc).

Based on regeneration data available at the time the SO_2-removal plant designs were made, a regeneration gas composition was selected in which the H_2 to CO ratio is fixed at about 7 or 8 to 1. The reformer pressure is set by pressure drop considerations (about 50 psi) and with the H_2/CO ratio also specified, reformer operating conditions can be calculated. Consequently it was found that a relatively high steam to carbon ratio (about 7 to 1) was required and the reformer outlet gas therefore contained a high percentage of water. While the acceptable concentration of water in the reducing gas feed to the regenerator has not been specified, it was assumed that it should be low (<10%) since water will probably lower the attrition resistance of the sorbent. A post-reforming treatment was included, therefore, whereby the water in the reformer is condensed and the "dry" gas reheated to the regenerator reaction temperature. Alternatively, a second reformer system was calculated in which the steam/carbon ratio is 1.4:1 and the concentration of water in the reformer effluent is about 8%. For this latter scheme it is assumed that no post-reforming treatment is needed. For purposes of identification, the high steam/carbon scheme is designated as the low temperature case while the low steam/carbon design is the high temperature case.

It should be noted that two basic assumptions are involved in the evaluation of the high temperature reformer design, viz,

(1) that a reforming catalyst is commercially available that will permit operation at the specified steam to carbon ratio; and (2) that reducing gas containing 8% water can be used to regenerate the sorbent without producing harmful side effects such as lowered attrition resistance. These assumptions naturally would have to be confirmed before a commercial plant incorporating these features could be built.

An alternative design which could be used would comprise using an intermediate steam to carbon ratio (i.e., between 7 and 1.4 to 1) in the reformer and adding a low temperature (say 700-900°F) shift converter downstream of the reformer. Thus the same final reducing gas composition could be achieved, possibly at a lower cost than required by the present base case design. The potential savings are probably small, however, and the net effect on the SO_2-removal plant costs probably would not be significant; therefore, the alternative design has not been evaluated since it was not believed to be justified for the present study. When and if an SO_2-removal plant is built, the reducing gas generation section would, of course, be optimized.

2. REGENERATION GAS COST

Cost of the regeneration gas produced by the steam-methane reformer is a function of:

- natural gas cost
- capital charge rate
- load factor
- degree of gas utilization (affects reformer size)
- SO_2-removal plant size (affects reformer size)
- steam price (excess steam is produced)

The effects on SO_2-removal plant costs have been determined for the variables listed above for each of the two reforming cases studied, i.e., the high and low temperature designs, and the results are presented in a series of graphs such that the operating cost of the reformer plant can be determined as dollars per ton of coal burned in the power plant. The reformer operating cost is obtained as follows:

1. For a 1000 MW plant size, 80% gas utilization, and 13% capital charge rate, operating cost, as cents per mole of reducing gas, is calculated for a range of natural gas and steam costs and load factors.

2. The reducing gas differential unit cost associated with variations in the gas utilization to the regenerator is obtained and added algebraically to the base cost found in item 1.

3. The reducing gas differential unit cost caused by variations in the capital charge rate (for a given plant size) is then added algebraically to the cost of items 1 and 2.

4. The unit cost of the reformed natural gas obtained from items 1, 2 and 3 is converted to a basis of dollars per ton of coal burned in the power plant.

The series of graphs developed for each case is used in the execution of the above steps as follows:

	REFORMER DESIGN	
STEP NUMBER	LOW TEMPERATURE	HIGH TEMPERATURE
1	Figure 33	Figure 39
2	Figure 34	Figure 40
3	Figure 34,36,37	Figure 41,42,43
4	Figure 38	Figure 44

FIGURE 33

COST OF REFORMED NATURAL GAS LOW TEMPERATURE CASE

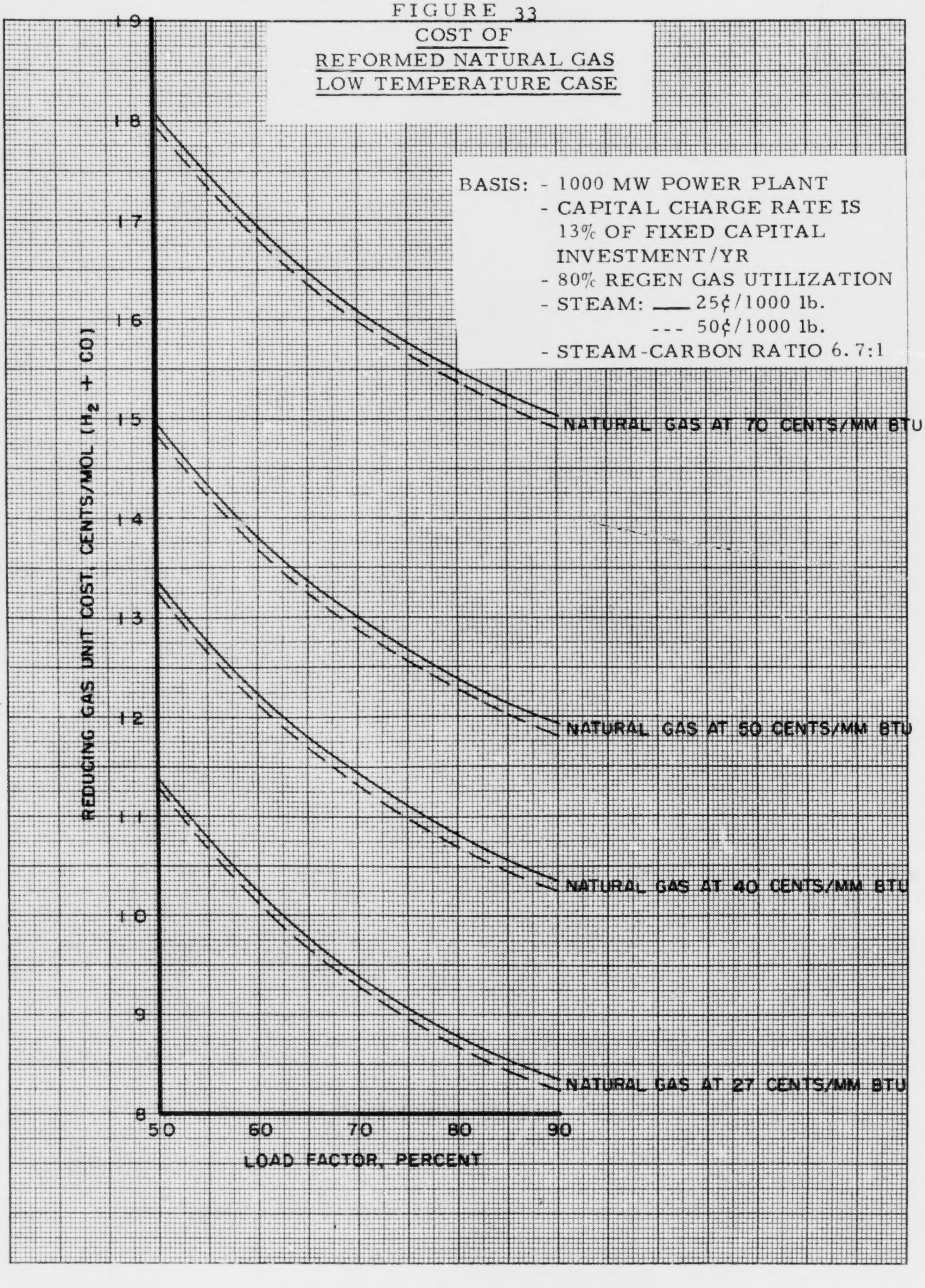
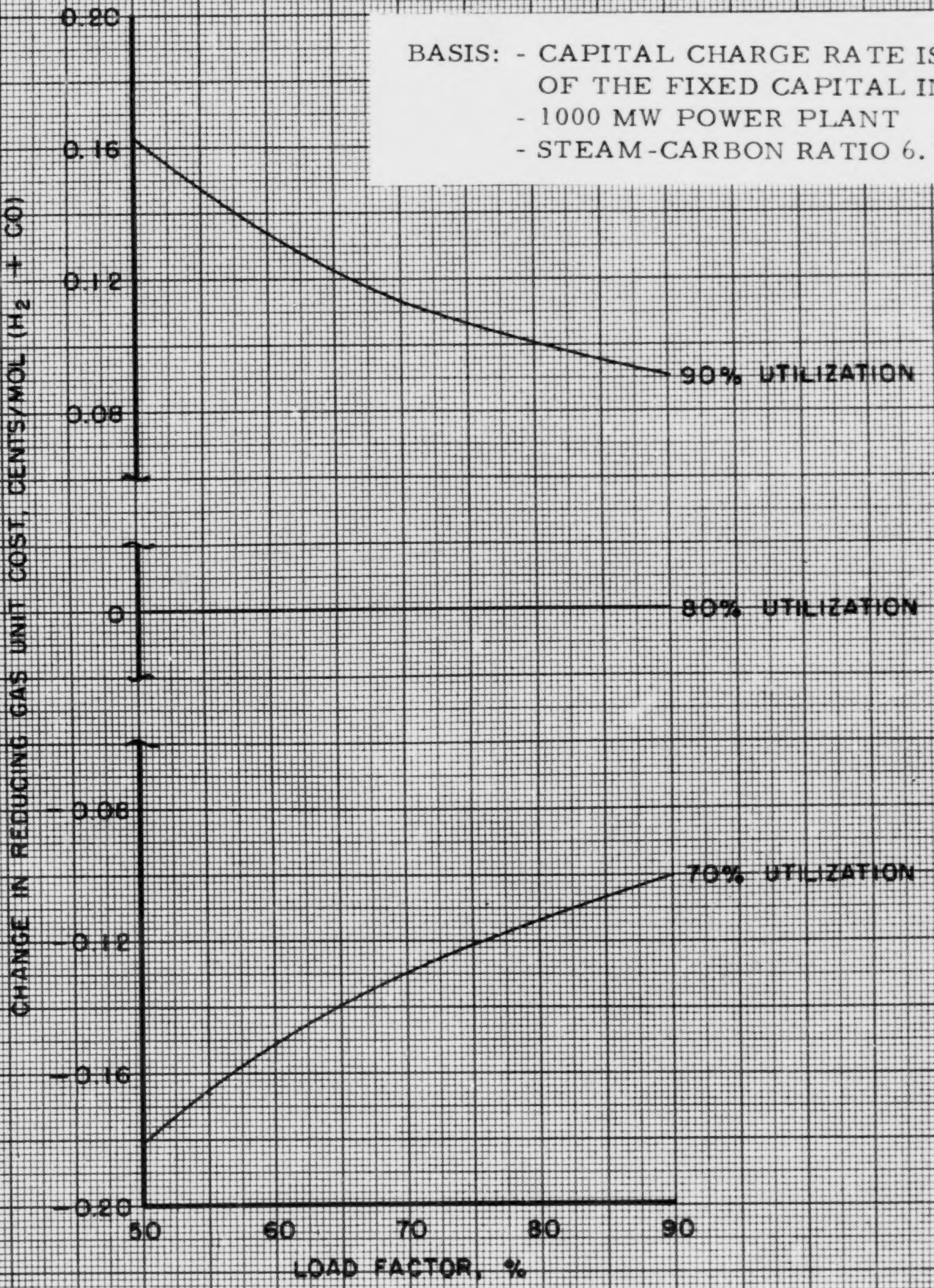


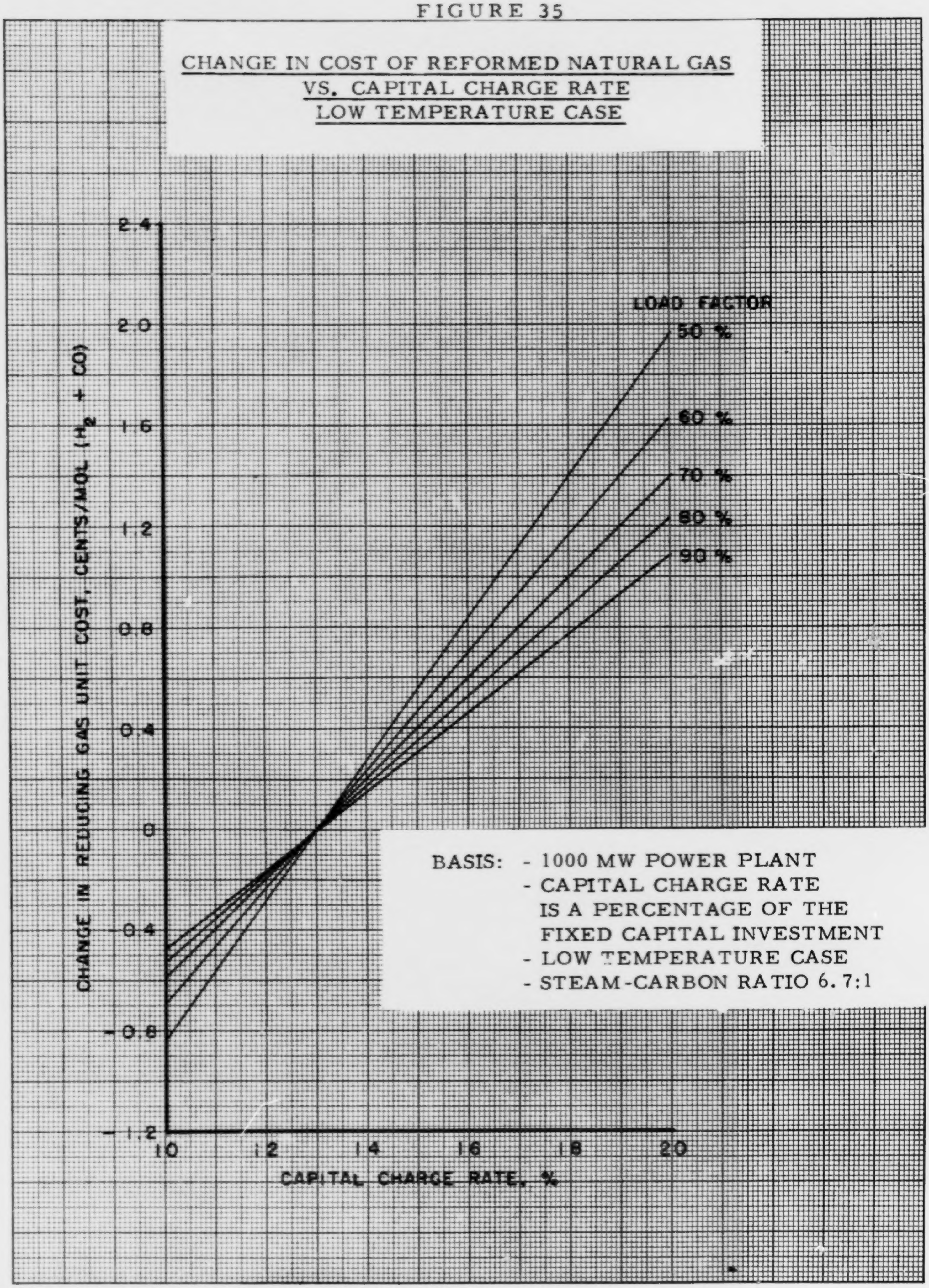
FIGURE 34

CHANGE IN COST OF REFORMED NATURAL GAS VS. GAS UTILIZATION IN THE REGENERATOR LOW TEMPERATURE CASE



BASIS: - CAPITAL CHARGE RATE IS 13% OF THE FIXED CAPITAL INVESTMENT
- 1000 MW POWER PLANT
- STEAM-CARBON RATIO 6.7:1

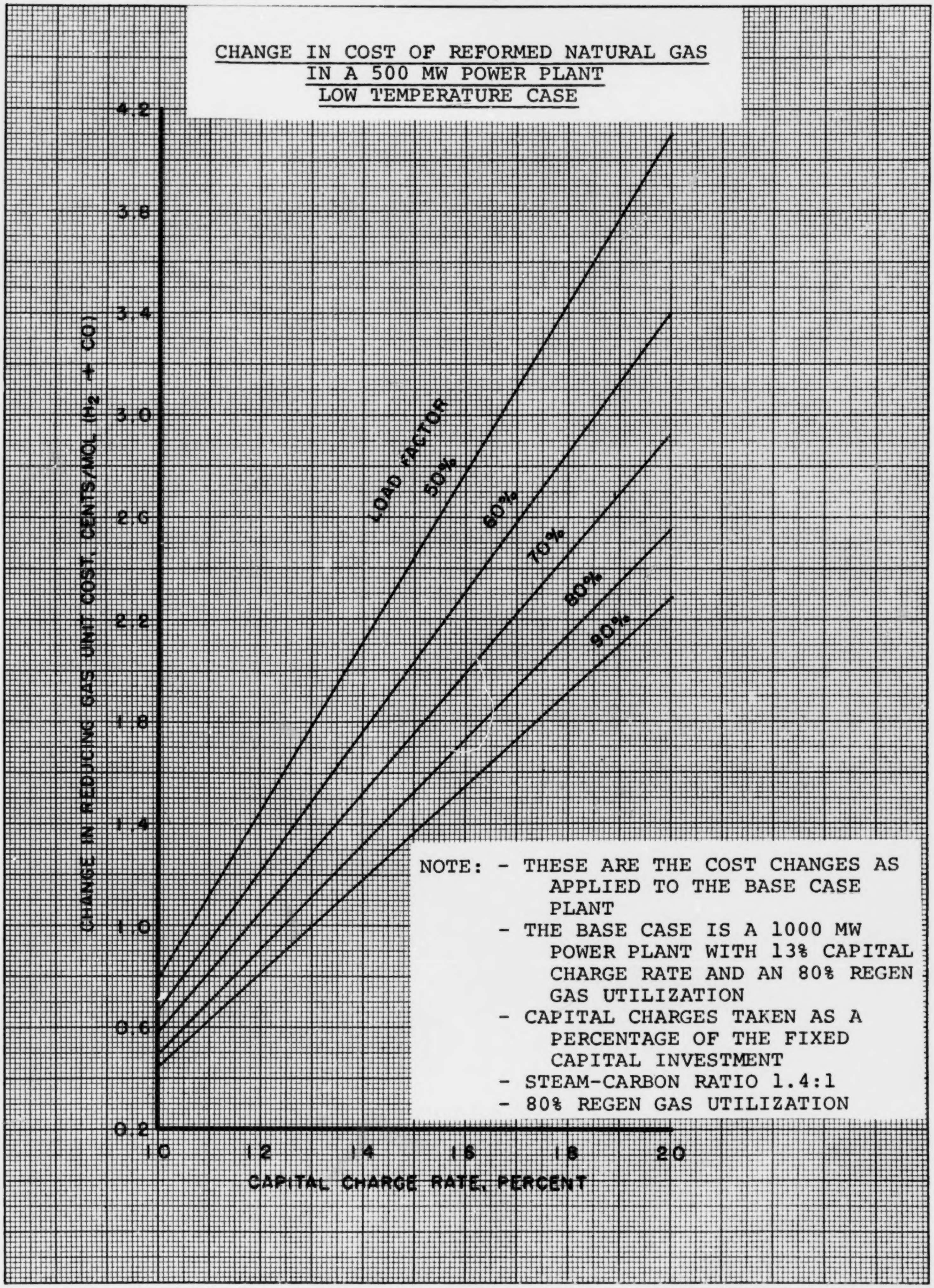
FIGURE 35



CHANGE IN COST OF REFORMED NATURAL GAS VS. CAPITAL CHARGE RATE LOW TEMPERATURE CASE

BASIS:
- 1000 MW POWER PLANT
- CAPITAL CHARGE RATE IS A PERCENTAGE OF THE FIXED CAPITAL INVESTMENT
- LOW TEMPERATURE CASE
- STEAM-CARBON RATIO 6.7:1

FIGURE 36

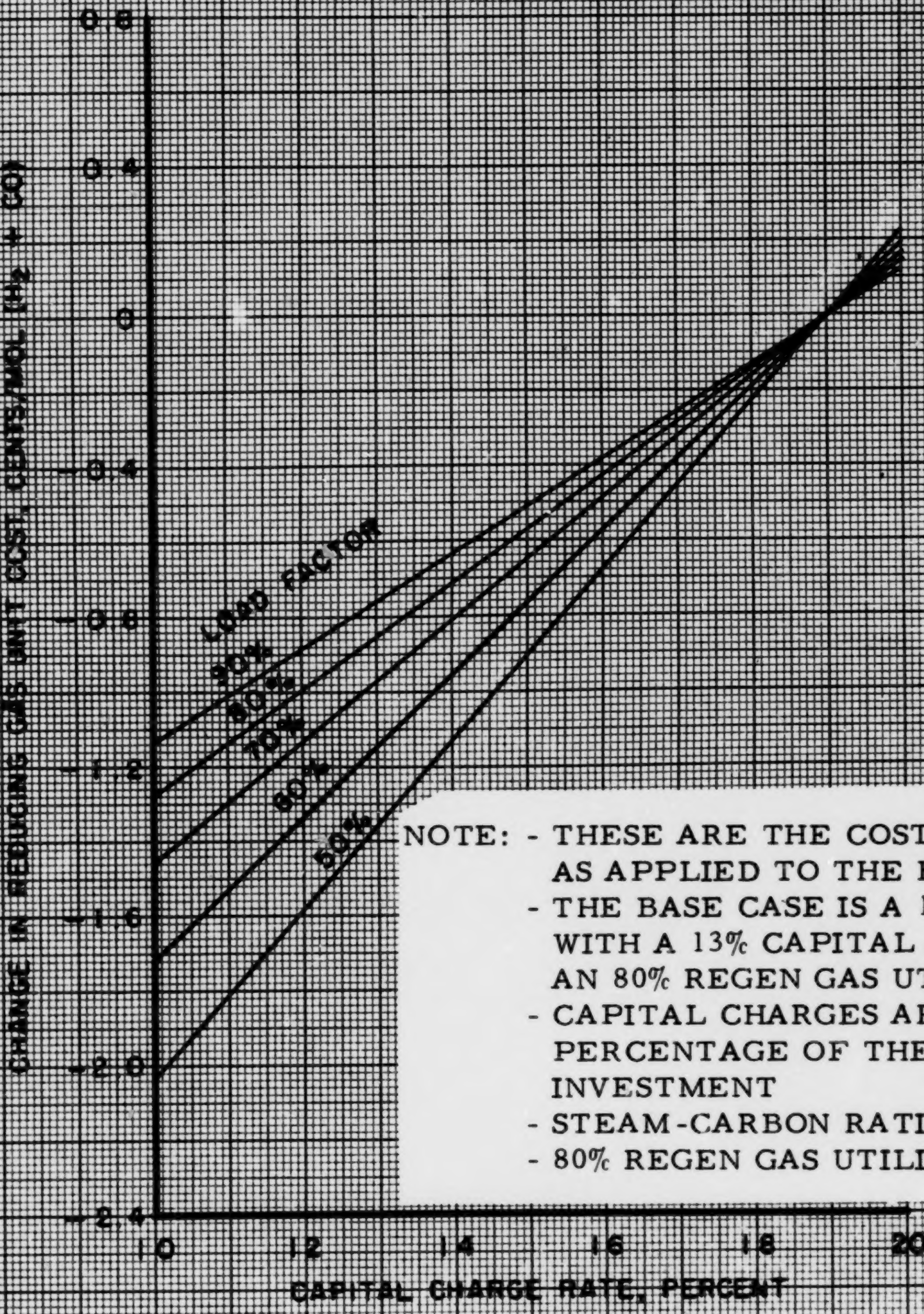


CHANGE IN COST OF REFORMED NATURAL GAS
IN A 500 MW POWER PLANT
LOW TEMPERATURE CASE

NOTE:
- THESE ARE THE COST CHANGES AS APPLIED TO THE BASE CASE PLANT
- THE BASE CASE IS A 1000 MW POWER PLANT WITH 13% CAPITAL CHARGE RATE AND AN 80% REGEN GAS UTILIZATION
- CAPITAL CHARGES TAKEN AS A PERCENTAGE OF THE FIXED CAPITAL INVESTMENT
- STEAM-CARBON RATIO 1.4:1
- 80% REGEN GAS UTILIZATION

FIGURE 37

CHANGE IN COST OF REFORMED NATURAL GAS
IN A 1500 MW POWER PLANT
LOW TEMPERATURE CASE



NOTE:
- THESE ARE THE COST CHANGES AS APPLIED TO THE BASE CASE PLANT
- THE BASE CASE IS A 1000 POWER PLANT WITH A 13% CAPITAL CHARGE RATE AND AN 80% REGEN GAS UTILIZATION
- CAPITAL CHARGES ARE TAKEN AS A PERCENTAGE OF THE FIXED CAPITAL INVESTMENT
- STEAM-CARBON RATIO 6.7:1
- 80% REGEN GAS UTILIZATION

FIGURE 38

UNIT COSTS CONVERSIONS
¢/MOL REDUCING GAS TO $/TON COAL BURNED
LOW TEMPERATURE CASE

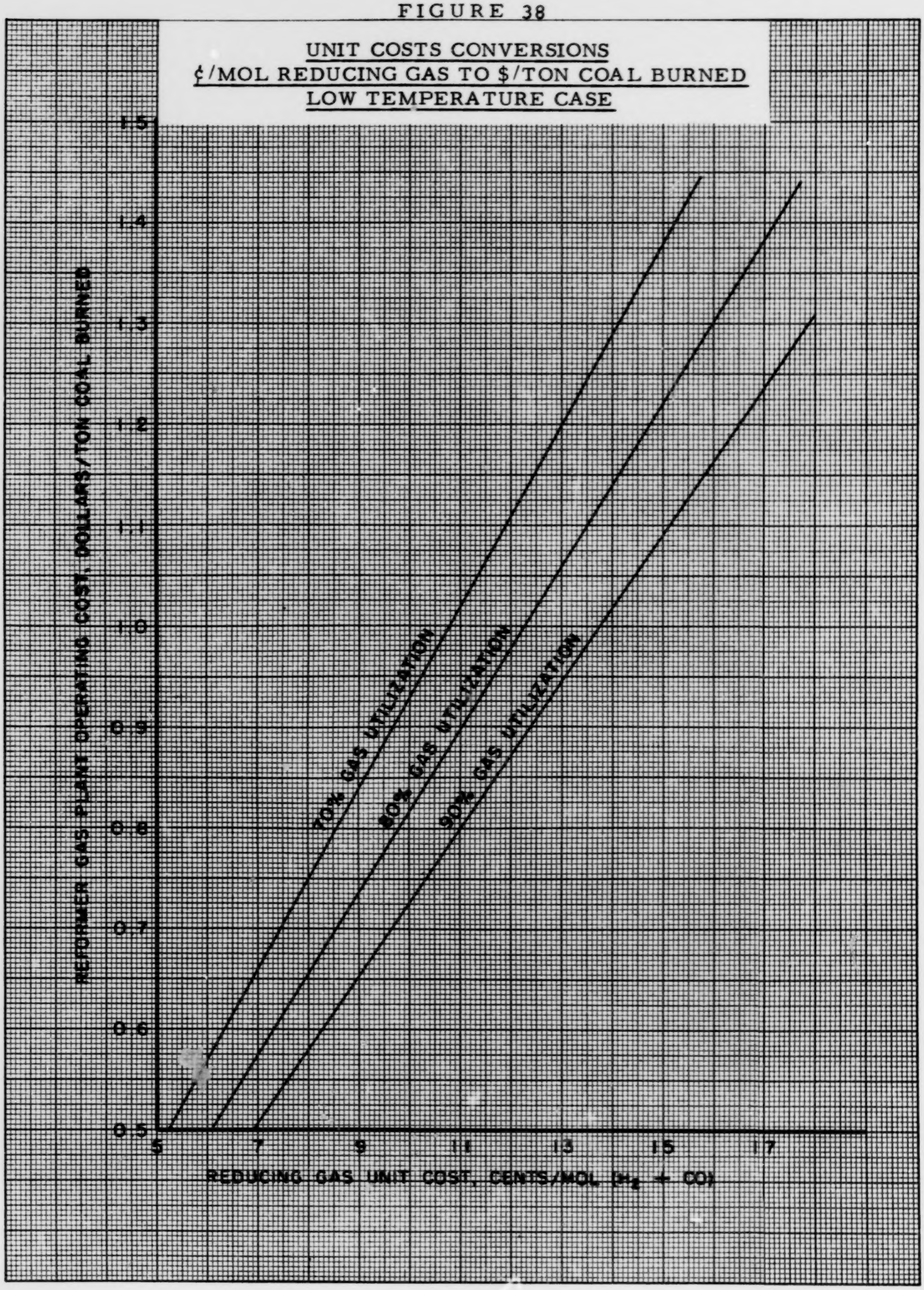
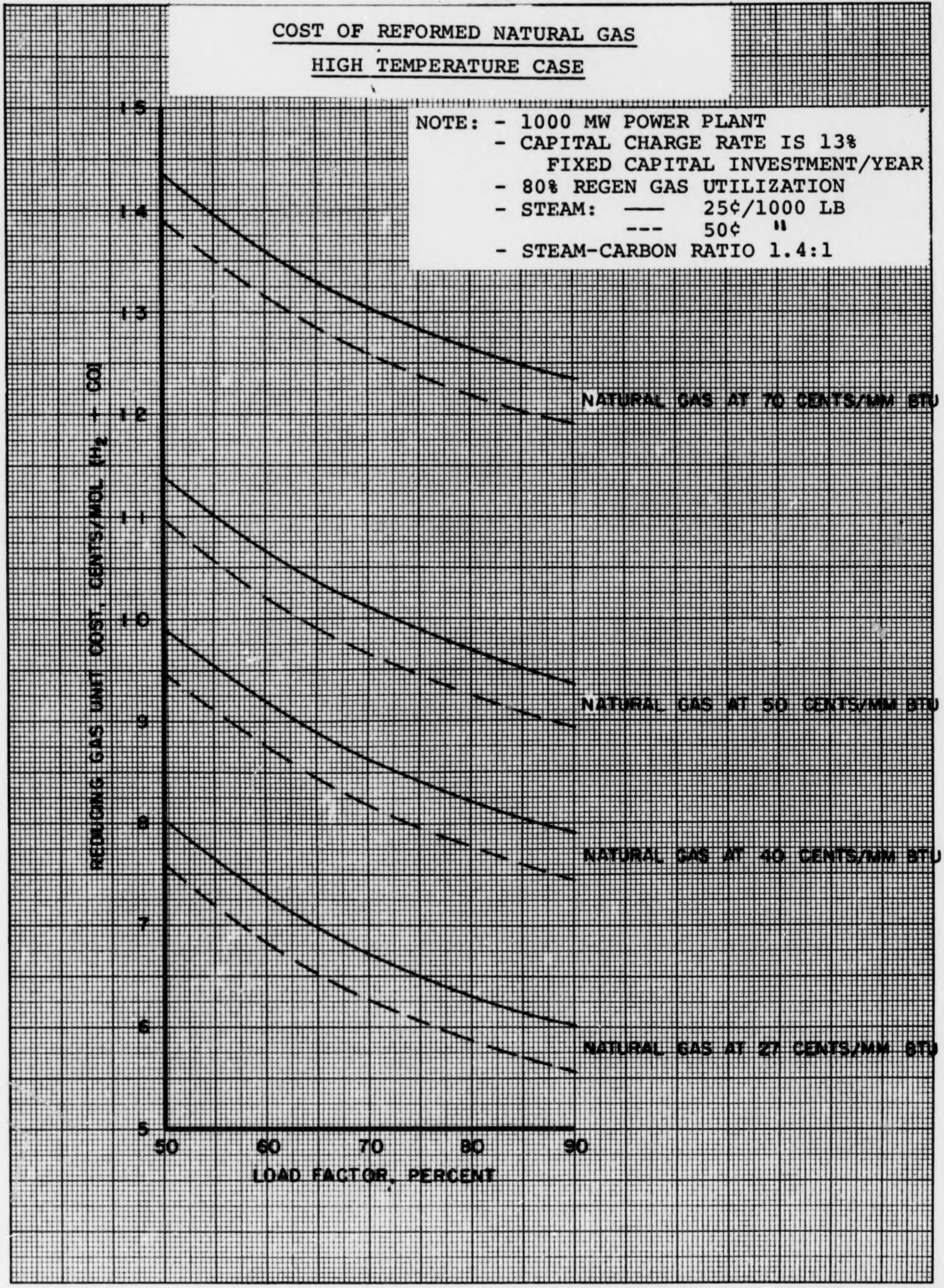


FIGURE 39



COST OF REFORMED NATURAL GAS

HIGH TEMPERATURE CASE

FIGURE 40

CHANGE IN COST OF REFORMED NATURAL GAS
VS. GAS UTILIZATION IN THE REGENERATOR
HIGH TEMPERATURE CASE

BASIS: - 1000 MW POWER PLANT
- CAPITAL CHARGE RATE IS 13% OF THE FIXED CAPITAL INVESTMENT
- STEAM-CARBON RATIO 1.4:1

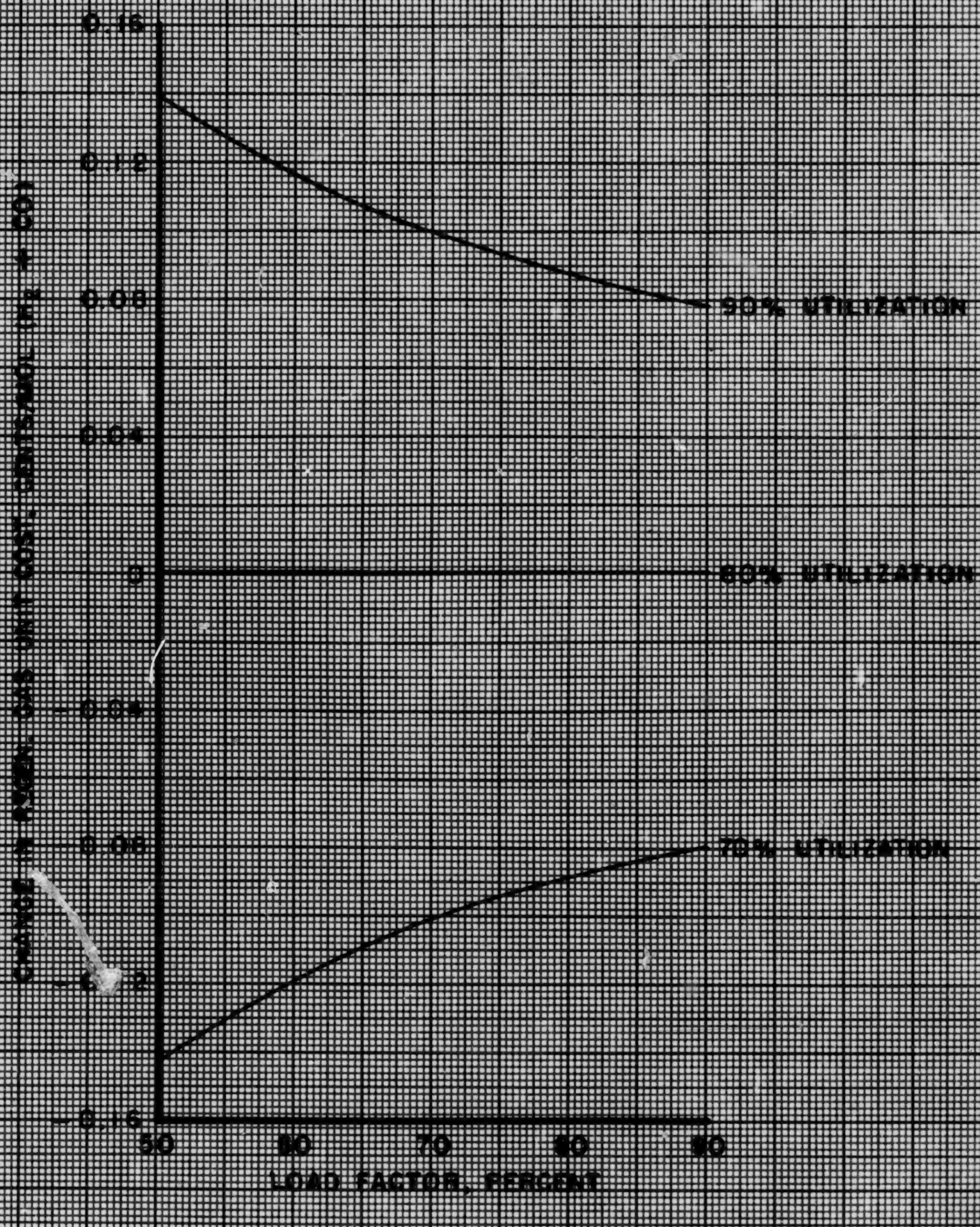
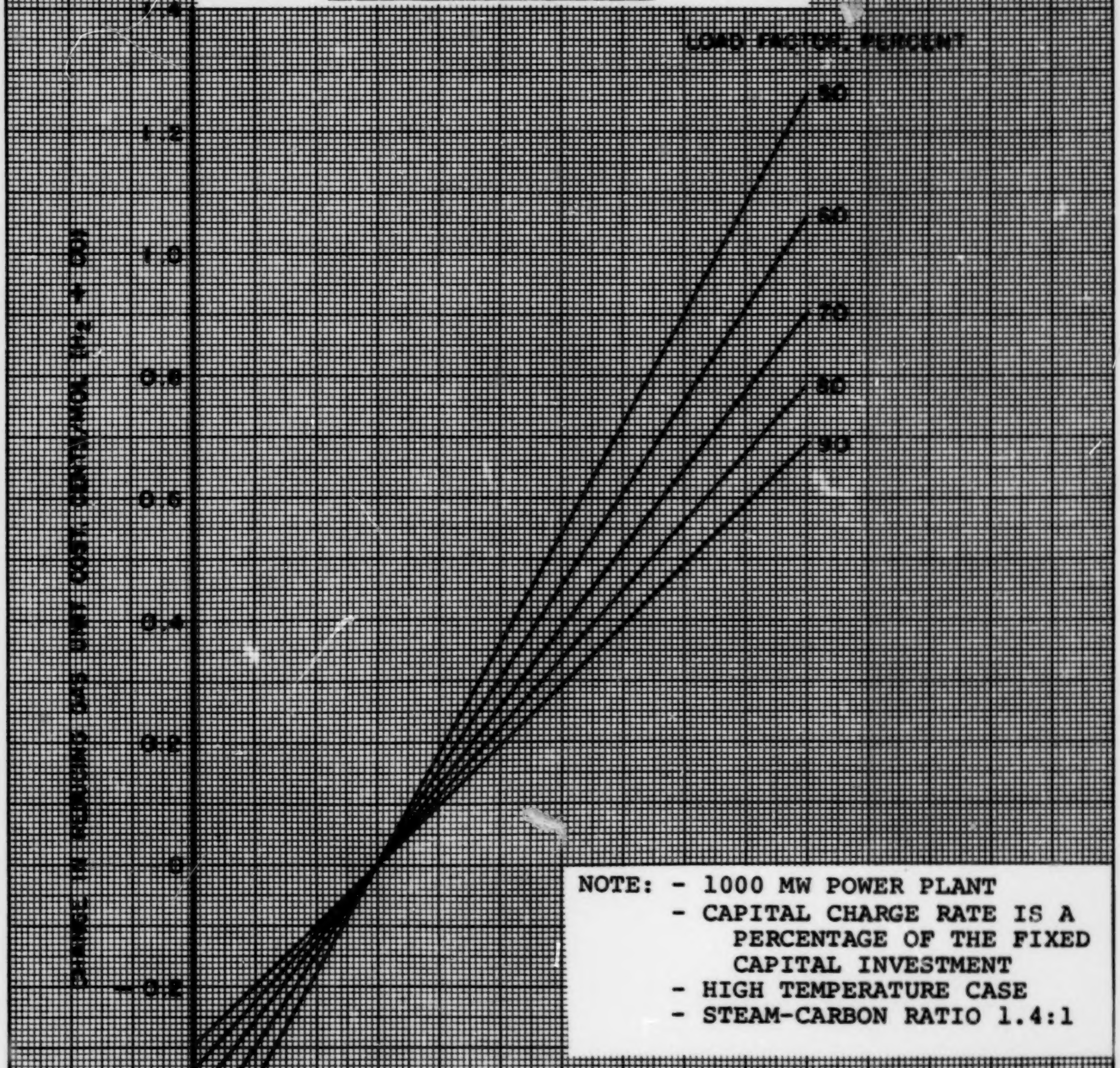


FIGURE 41

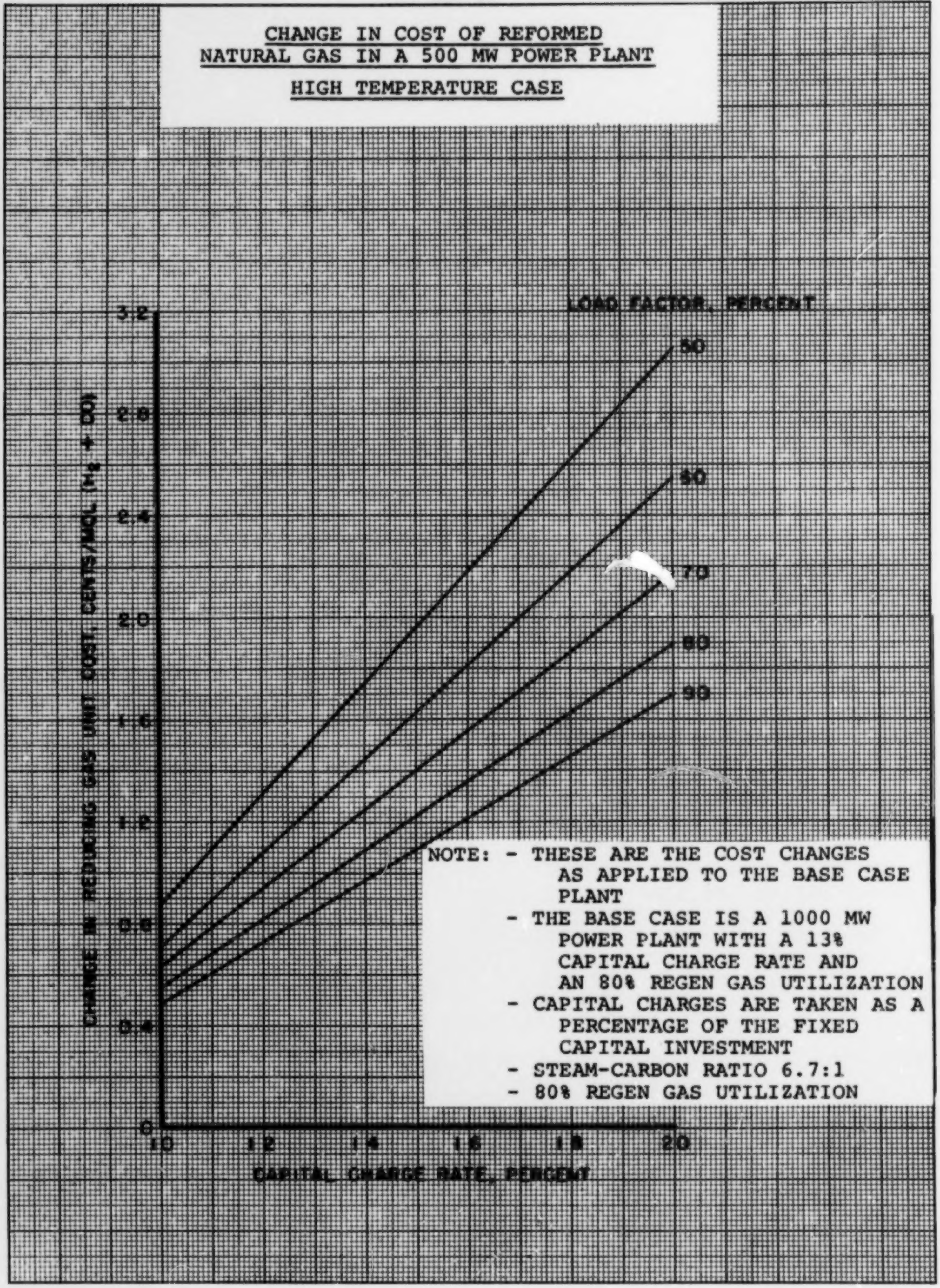
CHANGE IN COST OF REFORMED NATURAL GAS VS. THE CAPITAL CHARGE RATE

HIGH TEMPERATURE CASE



NOTE: - 1000 MW POWER PLANT
- CAPITAL CHARGE RATE IS A PERCENTAGE OF THE FIXED CAPITAL INVESTMENT
- HIGH TEMPERATURE CASE
- STEAM-CARBON RATIO 1.4:1

FIGURE 42

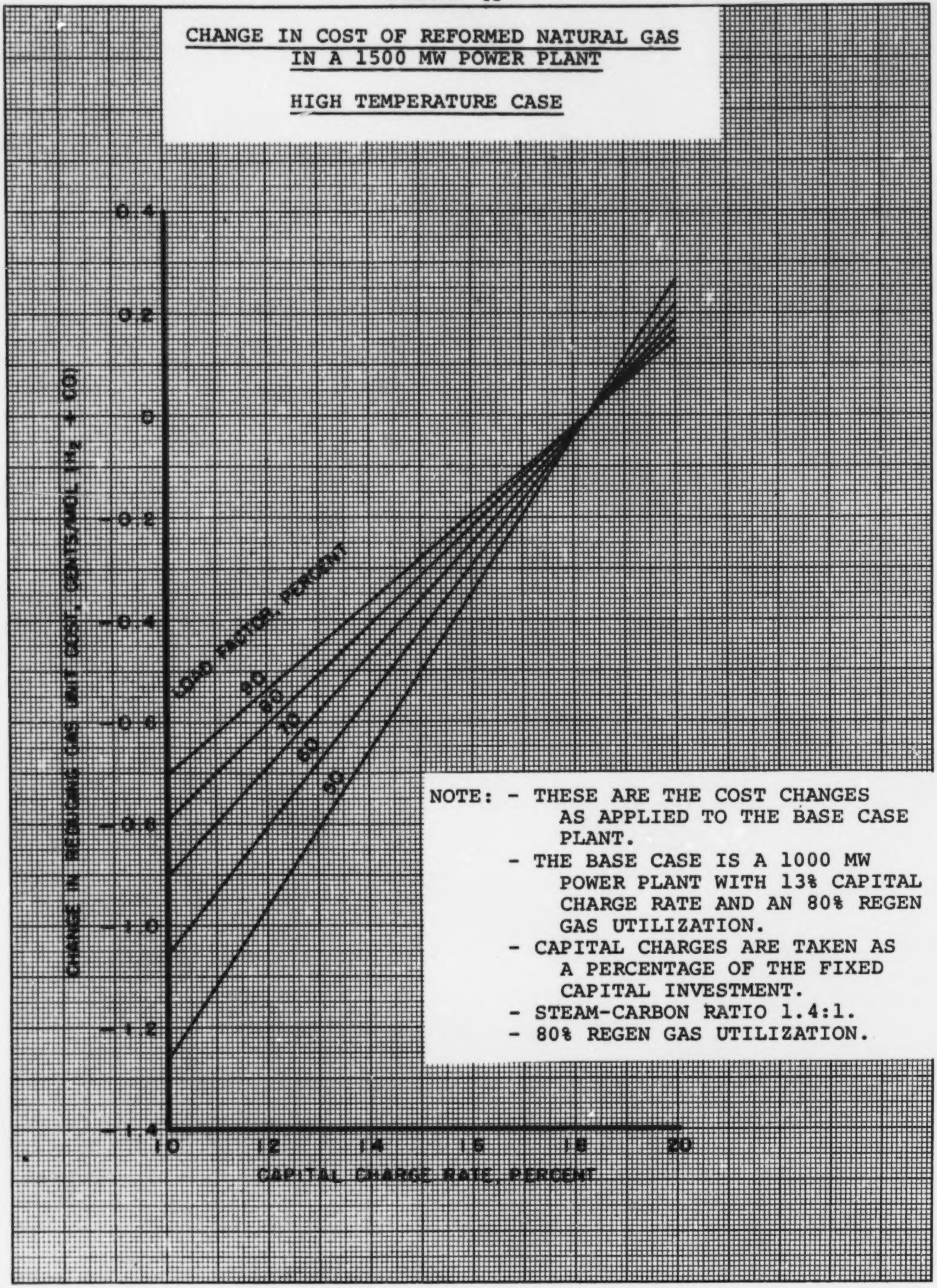


CHANGE IN COST OF REFORMED NATURAL GAS IN A 500 MW POWER PLANT

HIGH TEMPERATURE CASE

NOTE:
- THESE ARE THE COST CHANGES AS APPLIED TO THE BASE CASE PLANT
- THE BASE CASE IS A 1000 MW POWER PLANT WITH A 13% CAPITAL CHARGE RATE AND AN 80% REGEN GAS UTILIZATION
- CAPITAL CHARGES ARE TAKEN AS A PERCENTAGE OF THE FIXED CAPITAL INVESTMENT
- STEAM-CARBON RATIO 6.7:1
- 80% REGEN GAS UTILIZATION

FIGURE 43

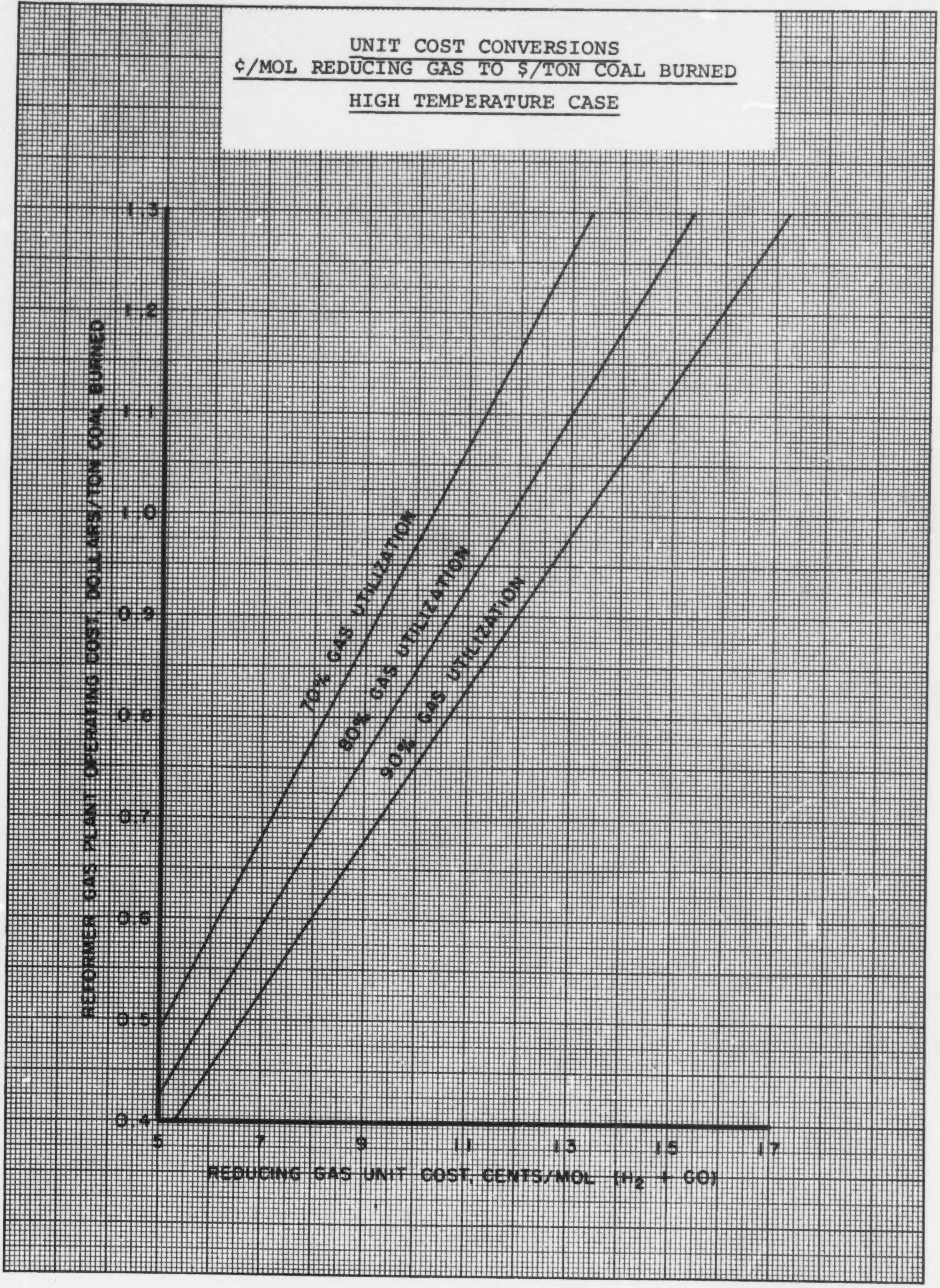


CHANGE IN COST OF REFORMED NATURAL GAS IN A 1500 MW POWER PLANT

HIGH TEMPERATURE CASE

NOTE:
- THESE ARE THE COST CHANGES AS APPLIED TO THE BASE CASE PLANT.
- THE BASE CASE IS A 1000 MW POWER PLANT WITH 13% CAPITAL CHARGE RATE AND AN 80% REGEN GAS UTILIZATION.
- CAPITAL CHARGES ARE TAKEN AS A PERCENTAGE OF THE FIXED CAPITAL INVESTMENT.
- STEAM-CARBON RATIO 1.4:1.
- 80% REGEN GAS UTILIZATION.

FIGURE 44



UNIT COST CONVERSIONS
¢/MOL REDUCING GAS TO $/TON COAL BURNED
HIGH TEMPERATURE CASE

From these the cost of the low temperature reforming is found to be $0.83/ton of coal at the base case conditions whereas the cost of high temperature reforming is $0.62/ton of coal. Referring to the base case design which incorporates the low temperature reformer, the reducing gas generation contributes 18.3%, 16.7%, and 15.0% to the operating costs of the dispersed phase, fluid bed, and fixed bed systems, respectively. If the low temperature reformer is replaced with the high temperature one, a cost reduction of $0.21/ton of coal is obtained, which is equivalent to 4-5% of the base case cost. Thus, if the high temperature reformer can be used, it would be preferred because of its lower operating costs.

Base case economics for both the high and low temperature reformers are based on the rather optimistic natural gas cost of 27¢/MM Btu. That is, the "summer price" has been assumed to apply all year. However, a natural gas cost of 50¢/MM Btu probably would better represent the annual average natural gas cost for a northern or north-central power plant location. If the latter value is used the annual operating costs become:

1.	Dispersed Phase	$4.98/Ton Coal
2.	Fluid Bed	$5.38/Ton Coal
3.	Fixed Bed	$5.95/Ton Coal

These figures represent an increase in operating cost of 9.9%, 8.0% and 7.8%, respectively and are based on a low temperature reformer. Operating cost of the low temperature reformer, itself, would be $1.12/ton coal at 50¢/MM Btu natural gas cost; therefore, its contribution to the total plant cost would be 22.5%, 20.8%, and 18.8% for the dispersed phase, fluid, and fixed bed schemes, respectively. If, however, the high temperature reformer were used, the total plant costs would be:

1.	Dispersed Phase	$4.76/Ton Coal
2.	Fluid Bed	$5.16/Ton Coal
3.	Fixed Bed	$5.73/Ton Coal

Operating cost of this high temperature reformer would be \$0.90/ton coal. This represents 18.9%, 17.4% and 15.7% of the total operating cost of the dispersed phase, fluid bed, and fixed bed sorption schemes, respectively. Thus, comparing the savings obtained by using the high temperature reformer, i.e., \$0.21/ton and \$0.22/ton at 27 and 50¢/MM Btu natural gas price respectively, it is seen that the reduction in costs obtained with the high temperature reformer is essentially independent of the natural gas price.

It should be noted that no differential is shown between the three base case reducing gas costs. While the dispersed phase and fluid bed processes do in fact require identical reformers, the fixed bed process reformer is slightly larger and the unit costs probably are somewhat less than for the other two cases. The difference is small, however, and for the present study has been neglected since it would have no significant effect on overall plant costs.

In the evaluation of reformer operating costs, the effect of utilization of the reducing gas in the regenerator was studied. Levels of utilization studied for each reformer were 70%, 80% (base case) and 90%. As can be seen from Figures 34 and 40 the differential cost due to gas utilization is small, the maximum change ranging from -0.18 to 0.16 cents/mol (H_2+CO) for the low temperature reformer, and from -0.14 to 0.14 cents/mol (H_2+CO) for the high temperature reformer. In terms of coal price this amounts to a range of from 0.11 \$/ton to 0.09\$/ton, which is about 1-2% of the operating cost for any gas utilization. Therefore, optimization of the reducing gas utilization in the regenerator will not significantly affect the total cost of the SO_2-removal plant.

To determine the effect of plant size on reformed natural gas costs, two additional plant sizes were examined and compared to the 1000 MW power plant base case. Plant sizes for 500 and

and 1500 MW power stations also were studied. The base case reformer cost for a 1000 MW power station is \$0.83/ton for the low temperature reformer, while in the 500 MW plant this cost increases 15.7% to \$0.96/ton, and in a 1500 MW plant this cost decreases 10.9% to \$0.74/ton. For the high temperature reformer the base case operating cost is \$0.62/ton, whereas in a 500 MW plant the operating cost is \$0.72/ton, a 16.1% increase. Similarly the 1500 MW power station's high temperature reformer would cost \$0.56/ton, a 9.7% decrease. Based on these values it appears that lowering the plant size by 50% increases the reducing gas cost by about 16% while a 50% larger plant shows gas costs that are 10% lower, the comparison basis being the 1000 MW plant.

A credit has been taken for excess steam produced in the reformer section. However, the value of this steam is not clearly established so two different prices have been used, viz, 25 and 50¢/M lbs. The lower value is used for the base case economics and is equal to the value of the fuel burned in the power plant. The higher value was chosen arbitrarily but it does represent a price that is commonly used for low pressure (~50 psig) steam cost.

The effect of the steam credit on reformer gas costs is shown on curves 33 and 39. Solid lines on these curves represent the 25¢/M lbs steam credit while the dashed lines are for a steam price of 50¢/M lbs. It is evident from the small difference between the solid and dashed lines that the price of by-product steam has little effect on SO_2-removal plant costs.

VII. SPECIFIC EXAMPLE OF USE OF CHARTS

To illustrate the use of the charts presented in Section VI, an example is given below based on process parameters selected arbitrarily but different from the base case conditions. A listing of the values selected for illustration is given in Table 25 along with those for the base case.

Since nearly all of the conditions refer to the base case which is for a 1000 MW power plant, to determine the costs for a different size plant when several variables are changed, as in the present example, the following procedure is used. First determine the new investment for the 1000 MW power plant and use it to derive the operating cost for the 1000 MW plant using the values specified for the various parameters. Next the operating cost is adjusted for plant size by applying the percentage correction as read from Figure 24 which shows the change in operating cost as a function of plant size between 500 and 1500 MW. Finally, the investment for the 1500 MW plant is calculated by adding the incremental cost for plant size which is obtained from Figure 22. It should be noted that this investment is not used in calculating operating costs in the present example.

The items (numbers correspond to those in Table 25) which affect investment are:

1. SO_2 removal efficiency
8. Percent sulfur in coal
9. Coal heating value/power plant heat rate
11. Sorbent loading

The adjusted investment for the 1000 MW plant is determined below:

TABLE 25

PROCESS SPECIFICATIONS -- BASE CASE AND ALTERNATIVE

		BASE CASE	ALTERNATE
1.	DESIGN (DP, FL B, FX B)*	DP	DP
2.	SO_2 REMOVAL EFF. %	90	70
3.	LOAD FACTOR %	60	80
4.	CAP. CHARGE RATE, % FCI	13	16
5.	MAINTENANCE, % FCI	5	4
6.	NAT. GAS COST, ¢/MM BTU	27	50
7.	STEAM CREDIT, ¢/M LB	25	50
8.	S IN COAL, %	3	6
9.	COAL HEATING VALUE, BTU/LB /POWER PLANT HEAT RATE, BTU/KWH	13,000/8600	10,000/8600
10.	PLANT SIZE, MW	1000	1500
11.	SORBENT LOADING, % S	3	6
12.	ATTRITION RATE, % SOLIDS THROUGHPUT /SORBENT PRICE, ¢/LB	0.1/25	0.2/25
13.	SULFUR CREDIT, $/TON	20	20
14.	REFORMED NAT. GAS	LOW TEMP	HIGH TEMP
	a) NAT. GAS COST, ¢/MM BTU	27	50
	b) UTILIZATION, %	80	90
	c) CAP. CHARGE RATE, % FCI	13	16
	d) PLANT SIZE, MW	1000	1500
15.	INVESTMENT, MM DOLLARS	30,720,000	-
16.	UNIT COSTS, $/TON COAL	4.53	-

* DP - DISPERSED PHASE
FL B- FLUID BED
FX B- FIXED BED

ITEM	INVESTMENT, $	FIGURE NO.
BASE CASE (DISPERSED PHASE	30,720,000	-
1. 70% REMOVAL	- 2,800,000	1
8. 6% S COAL	+ 9,000,000	17
*9. 10,000 BTU/LB/8600 BTU/KWH	+ 1,280,000	(SEE BELOW)
11. 6% LOADING	- 260,000	27
TOTAL ADJUSTED INVESTMENT =	37,940,000	

*Calculation of investment change attributable to coal heating value and heat rate:

The base case flue gas rate is 288.6 M mols/hr and the volume is 3.58 MM ACFM; these values are marked on Figure 20. The flue gas rate for 10,000 Btu/lb and 8600 Btu/kwh is read from Figure 19 and is 305 M Mols/hr. Since the pressure and temperature are unchanged, the gas volume can be calculated from the moles directly, thus: $\frac{305}{288.6} X 3.58 = 3.78$MM ACFM. From Figure 20 the total investment at 3.78 MM ACFM is found to be $32,000,000. However, since the change in investment is desired, it is obtained by subtracting the base case value: $32,000,000 -$30,720,000 = $1,280,000.

The operating costs for the 1000 MW plant are calculated in Table 26 using the adjusted investment of $37,940,000 and the values listed in Table 25. The cost of the 1000 MW plant calculated in Table 26 is now adjusted for plant size. From Figure 24 the change for a 1500 MW plant is -14% and the 1000 MW cost of $4.64/ton is therefore reduced by this amount and becomes $3.99/ton.

The investment for the 1500 MW plant is obtained by adding the incremental cost from Figure 22, viz, $9,200,000. Thus for the dispersed phase design the total plant investment for the specified condition is $47,140,000 (37,940,000 + 9,200,000).

Other combinations can be evaluated in a similar manner. Although all of the charts were not used in the above illustration, the procedure is the same and therefore is not shown here.

TABLE 26

SAMPLE CALCULATION OF UNIT COST

1000 MW PLANT

	ITEM	UNIT COST $/TON	FIGURE NO.
	BASE CASE	4.53	-
2.	70% REMOVAL	-0.42	2
3.	80% LOAD FACTOR	-0.35	5
[1] 4.	16% CAPITAL CHARGE RATE	+0.50	9,10
[2] 5.	4% MAINTENANCE	-0.38	11,12
6.	NAT. GAS @ 50¢/MM BTU	+0.45	13
7.	STEAM CREDIT, 50¢/M LB	-0.06	14
8.	6% S IN COAL	+1.53	18
[3] 9.	10,000 BTU/# COAL & 8600 BTU/KWH	-0.10	21
11.,12.	6% LOADING, 0.2% ATTRITION	-0.15	28
[4] 13.	SULFUR AT $20/TON	-0.84	p.94
[5] 14.	REDUCING GAS (HIGH TEMP)	-0.07	
	TOTAL	4.64	

[1] Y = 0.165, Fig. 9; at Y = 0.165, Δ Cost = +0.50 Fig. 10.

[2] @ 4%, Cost = 1.51, Fig. 11; using 1.51, Total Cost = 1.15 Fig. 12. However, Δ cost is needed and since base case cost is 1.53 which is indicated on Fig. 11, Δ = 1.15-1.53 = -0.38.

[3] Calculations under Item 9 for investment changes show flue gas volume = 3.78 MM ACFM. From Fig. 21 the unit cost is 4.43 and Δ cost = 4.43-4.53 = -0.10.

[4] The sulfur credit for 6% sulfur is twice that for 3% S or 2X0.54 = $1.08/ton. However, this credit must be adjusted for 70% removal since the base case is for 90% removal, thus: 1.08X0.7/0.9 = 0.84.

TABLE 26

(CONTINUED)

[5] Base case reducing gas cost at 27¢/MM Btu and 25¢/M lb steam is $0.83/ton coal from Figures 33 and 38. This cost is adjusted to high temperature, 50¢/MM Btu nat. gas, 50¢/M lb steam, and 80% load factor thus:

Basic Cost = 9.7¢/Mol Fig. 39
90% Utilization = +0.088¢/Mol Fig. 40
16% Cap. Charge = +0.34¢/Mol Fig. 41
Total 10.128
@ 10.128¢/Mol, Cost = $0.76/ton, Fig. 44
and Δ cost = 0.76-0.83 = -0.07.

VIII. COST CALCULATION METHOD -- COMPUTER PROGRAM

The calculation method for the annual and unit operating costs for the present study is summarized in Tables 27 - 29, which are computer printouts of the base cases. The computer program which was developed to perform these calculations is shown following Tables 27 - 29.

TABLE 28

```
                                   FLUID BED

                                 STANDARD CASE

                       PROCESS ECONOMICS  -  COMPLETE PLANT


                FIXED CAPITAL INVESTMENT  ( FCI )      $ 35900008.
                STREAM TIME      HOURS PER YEAR              5250.


DIRECT COST
                                                                              $ / YEAR

    1.  OPERATING LABOR ( 5.00 MEN / SHIFT AT $3.50 / HR )                $    153300.
    2.  SUPERVISION - 15 PERCENT OF ITEM 1                                      22995.
    3.  MAINTENANCE LABOR - 3 PERCENT OF FCI                                  1077000.
    4.  MAINTENANCE MATERIAL - 2 PERCENT OF FCI                                718000.
    5.  SUPPLIES - 15 PFRCENT OF ITEMS ( 3 + 4 )                               269249.
    6.  UTILITIES ,
           A.  COOLING WATER - 5020. GPM AT    5.0 CENTS / M GAL.               79065.
           B.  NATURAL GAS - 570.519 MM BTU / HOUR AT  27.0 CENTS / MM BTU     808711.
           C.  POWER - 21788. KW AT 0.60 CENTS / KWH                           686341.
    7.  CREDITS ,
           A.  HEAT - 153.114 MM BTU / HOUR AT 25.0 CENTS / MM BTU         (   200962.)
           B.  STEAM -  77360. POUNDS / HOUR AT  25.0 CENTS / M LBS.       (   101535.)

    8.  SUBTOTAL DIRECT COST   - - - - - - - - - - - - - - - - - - - - -  $   3512164.

    9.  SULFUR CREDIT -  17855. POUNDS / HOUR AT  20.00 $ / TON           (   937387.)
   10.  ATTRITION LOSS - 0.050 PERCENT OF THE   379800. POUNDS / HOUR          249243.
                           CIRCULATION RATE AT 0.25 $ / POUND
   11.  TOTAL DIRECT COST    - - - - - - - - - - - - - - - - - - - - - -  $   2824020.
INDIRECT COST

   12.  PAYROLL OVERHEAD - 20 PERCENT OF ( 1 + 2 )                              35259.
   13.  PLANT OVERHEAD - 50 PERCENT OF ( 1 + 2 + 3 + 4 + 5 )                  1120272.
   14.  TOTAL INDIRECT COST   - - - - - - - - - - - - - - - - - - - - - -  $   1155531.
FIXED COST

   15.  CAPITAL CHARGES - 13.00 PERCENT OF FCI.                            $   4667000.
           ( INCLUDES DEPRECIATION, INSURANCE, TAXES, COST OF CAPITAL, ETC. )


TOTAL OPERATING COST

   16.  TOTAL OPERATING COST - ITEMS ( 11 + 14 + 15 ) - - - - - - - - - -  $   8646550.
UNIT COST

   17.  $ / TON COAL AT  330.768 TONS / HOUR                                     4.979
   18.  MILS / KWH AT 1000 MW CAPACITY                                           1.646
```

TABLE 27

DISPERSED PHASE

STANDARD CASE

PROCESS ECONOMICS - COMPLETE PLANT

```
                     FIXED CAPITAL INVESTMENT  ( FCI )     $ 30720004.
                     STREAM TIME     HOURS PER YEAR             5250.


DIRECT COST                                                                         $ / YEAR

    1.  OPERATING LABOR ( 4.00 MEN / SHIFT AT $3.50 / HR )                      $    122640.
    2.  SUPERVISION - 15 PERCENT OF ITEM 1                                            18396.
    3.  MAINTENANCE LABOR - 3 PERCENT OF FCI                                         921600.
    4.  MAINTENANCE MATERIAL - 2 PERCENT OF FCI                                      614400.
    5.  SUPPLIES - 15 PERCENT OF ITEMS ( 3 + 4 )                                     230399.
    6.  UTILITIES ,
           A.  COOLING WATER - 5020. GPM AT   5.0 CENTS / M GAL.                      79065.
           B.  NATURAL GAS - 649.796 MM BTU / HOUR AT  27.0 CENTS / MM BTU           921085.
           C.  POWER - 12239. KW AT 0.60 CENTS / KWH                                 385533.
    7.  CREDITS ,
           A.  HEAT - 223.232 MM BTU / HOUR AT 25.0 CENTS / MM BTU              (    292992.)
           B.  STEAM -  77360. POUNDS / HOUR AT  25.0 CENTS / M LBS.            (    101535.)

    8.  SUBTOTAL DIRECT COST    - - - - - - - - - - - - - - - - - - - - - - -   $   2898593.

    9.  SULFUR CREDIT -  17855. POUNDS / HOUR AT  20.00 $ / TON                 (    937387.)
   10.  ATTRITION LOSS - 0.100 PERCENT OF THE   705792. POUNDS / HOUR                926352.
                                  CIRCULATION RATE AT 0.25 $ / POUND
   11.  TOTAL DIRECT COST   - - - - - - - - - - - - - - - - - - - - - - - - -   $   2887557.

 INDIRECT COST

   12.  PAYROLL OVERHEAD - 20 PERCENT OF ( 1 + 2 )                                    28207.
   13.  PLANT OVERHEAD - 50 PERCENT OF ( 1 + 2 + 3 + 4 + 5 )                         953717.
   14.  TOTAL INDIRECT COST    - - - - - - - - - - - - - - - - - - - - - - - -   $    981925.

 FIXED COST

   15.  CAPITAL CHARGES - 13.00 PERCENT OF FCI,                                 $   3993600.
            ( INCLUDES DEPRECIATION, INSURANCE, TAXES, COST OF CAPITAL, ETC. )


 TOTAL OPERATING COST

   16.  TOTAL OPERATING COST - ITEMS ( 11 + 14 + 15 ) - - - - - - - - - - - - -  $   7863082.

 UNIT COST

   17.  $ / TON COAL AT  330.768 TONS / HOUR                                          4.528
   18.  MILS / KWH AT 1000 MW CAPACITY                                                1.497
```

TABLE 28

FLUID BED

STANDARD CASE

PROCESS ECONOMICS - COMPLETE PLANT

```
                         FIXED CAPITAL INVESTMENT  ( FCI )      $ 35900008.
                         STREAM TIME      HOURS PER YEAR              5250.


DIRECT COST                                                                                  $ / YEAR

   1.  OPERATING LABOR ( 5.00 MEN / SHIFT AT $3.50 / HR )                               $     153300.
   2.  SUPERVISION - 15 PERCENT OF ITEM 1                                                      22995.
   3.  MAINTENANCE LABOR - 3 PERCENT OF FCI                                                  1077000.
   4.  MAINTENANCE MATERIAL - 2 PERCENT OF FCI                                                718000.
   5.  SUPPLIES - 15 PERCENT OF ITEMS ( 3 + 4 )                                               269249.
   6.  UTILITIES ,
           A.  COOLING WATER - 5020. GPM AT    5.0 CENTS / M GAL.                              79065.
           B.  NATURAL GAS - 570.519 MM BTU / HOUR AT  27.0 CENTS / MM BTU                    808711.
           C.  POWER - 21788. KW AT 0.60 CENTS / KWH                                          686341.
   7.  CREDITS ,
           A.  HEAT - 153.114 MM BTU / HOUR AT 25.0 CENTS / MM BTU                      (     200962.)
           B.  STEAM -  77360. POUNDS / HOUR AT  25.0 CENTS / M LBS.                    (     101535.)

   8.  SUBTOTAL DIRECT COST   - - - - - - - - - - - - - - - - - - - - - - - - - - - -   $    3512164.

   9.  SULFUR CREDIT -  17855. POUNDS / HOUR AT  20.00 $ / TON                          (     937387.)
  10.  ATTRITION LOSS - 0.050 PERCENT OF THE    379800. POUNDS / HOUR                         249243.
                                  CIRCULATION RATE AT 0.25 $ / POUND
  11.  TOTAL DIRECT COST   - - - - - - - - - - - - - - - - - - - - - - - - - - - - - -   $    2824020.

INDIRECT COST

  12.  PAYROLL OVERHEAD - 20 PERCENT OF ( 1 + 2 )                                               35259.
  13.  PLANT OVERHEAD - 50 PERCENT OF ( 1 + 2 + 3 + 4 + 5 )                                  1120272.
  14.  TOTAL INDIRECT COST   - - - - - - - - - - - - - - - - - - - - - - - - - - - - -   $    1155531.

FIXED COST

  15.  CAPITAL CHARGES - 13.00 PERCENT OF FCI,                                          $    4667000.
           ( INCLUDES DEPRECIATION, INSURANCE, TAXES, COST OF CAPITAL, ETC. )


TOTAL OPERATING COST

  16.  TOTAL OPERATING COST - ITEMS ( 11 + 14 + 15 ) - - - - - - - - - - - - - - - - - -   $    8646550.

UNIT COST

  17.  $ / TON COAL AT  330.768 TONS / HOUR                                                     4.979
  18.  MILS / KWH AT 1000 MW CAPACITY                                                           1.646
```

TABLE 29

```
                                FIXED BED

                              STANDARD CASE

                    PROCESS ECONOMICS  -  COMPLETE PLANT


          FIXED CAPITAL INVESTMENT  ( FCI )      $ 40210008.
          STREAM TIME     HOURS PER YEAR               5250.


                                                                                  $ / YEAR
DIRECT COST

   1.  OPERATING LABOR ( 5.00 MEN / SHIFT AT $3.50 / HR )                      $    153300.
   2.  SUPERVISION - 15 PERCENT OF ITEM 1                                            22995.
   3.  MAINTENANCE LABOR - 3 PERCENT OF FCI                                        1206300.
   4.  MAINTENANCE MATERIAL - 2 PERCENT OF FCI                                      804200.
   5.  SUPPLIES - 15 PERCENT OF ITEMS ( 3 + 4 )                                     301574.
   6.  UTILITIES ,
         A.  COOLING WATER - 8773. GPM AT   5.0 CENTS / M GAL.                      138180.
         B.  NATURAL GAS - 618.288 MM BTU / HOUR AT  27.0 CENTS / MM BTU            876424.
         C.  POWER - 22170. KW AT 0.60 CENTS / KWH                                  698367.
   7.  CREDITS ,
         A.  HEAT - 129.170 MM BTU / HOUR AT 25.0 CENTS / MM BTU               (    169535.)
         B.  STEAM -  84140. POUNDS / HOUR AT  25.0 CENTS / M LBS.             (    110433.)

   8.  SUBTOTAL DIRECT COST    - - - - - - - - - - - - - - - - - - - - - - -   $   3921372.

   9.  SULFUR CREDIT -  19189. POUNDS / HOUR AT  20.00 $ / TON                 (   1007422.)
  10.  ATTRITION LOSS - 3.048 YEARS / INVENTORY LOSS AT 0.25 $ / POUND              159120.
                             AND A   1940000. POUND INVENTORY
  11.  TOTAL DIRECT COST   - - - - - - - - - - - - - - - - - - - - - - - - -   $   3073070.

 INDIRECT COST

  12.  PAYROLL OVERHEAD - 20 PERCENT OF ( 1 + 2 )                                    35259.
  13.  PLANT OVERHEAD - 50 PERCENT OF ( 1 + 2 + 3 + 4 + 5 )                        1244184.
  14.  TOTAL INDIRECT COST   - - - - - - - - - - - - - - - - - - - - - - - -   $   1279443.

 FIXED COST

  15.  CAPITAL CHARGES - 13.00 PERCENT OF FCI,                                 $   5227300.
         ( INCLUDES DEPRECIATION, INSURANCE, TAXES, COST OF CAPITAL, ETC. )


 TOTAL OPERATING COST

  16.  TOTAL OPERATING COST - ITEMS ( 11 + 14 + 15 ) - - - - - - - - - - - -   $   9579812.

 UNIT COST

  17.  $ / TON COAL AT  330.768 TONS / HOUR                                          5.516
  18.  MILS / KWH AT 1000 MW CAPACITY                                                1.824
```

TABLE 30

ECONOMIC VARIABLES INCLUDED IN THE COMPUTER PROGRAM FOR PHASE 3 EVALUATIONS

1. Number of men per shift
2. Sorbent cost
3. Cooling water cost
4. Natural gas cost
5. Power cost
6. Heat credit
7. Steam credit
8. Capital charges as a percentage of fixed capital investment
9. Fixed capital investment less the cost of the sorbent inventory
10. Steam factor
11. Sorbent inventory
12. Sorbent circulation rate
13. Sulfur production
14. Cooling water flow
15. Power requirement
16. Steam production
17. Sulfur price
18. Sorbent attrition rate
19. Power plant coal consumption
20. Heat credit
21. Natural gas consumption

```
// JOB T
// FOR
*NAME                    ECHO
* IOCS(CARD,TYPEWRITER,1132 PRINTER)
* ARITHMETIC TRACE
* LIST ALL
* TRANSFER TRACE
* ONE WORD INTEGERS
      DIMENSION VAR(21,10),M(21),N(21)
      COMMON VAR, M, N, FCI, A1,A2,A3,A4,A5,A6A,A6A1,A6A2,A6B,A6B1,A6C
      COMMON A6C1,A6C2,A7A,A7A1,A7B,A7B1,A8,A9,A10,A11,A12,A13,A14,A15
      COMMON A16,A17,A18,K1,K2,K3,K4,K5,K6,K7,K8,K9,K10,K11,K12,K13,K14
      COMMON K15,K16,K17,K18,K19,K20,K21,IR,IW,INDEX,KEY,LOCK
      IR = 2
      IW = 3
      READ (IR,100) (N(I),I = 1,21 )
  100 FORMAT(8I10)
      DO 1 I = 1,21
      DO 1 J = 1,10
      M(I) = 1
    1 VAR(I,J) = 0.0
C     THE FOLLOWING SETS OF CONSTANTS ARE TO BE USED AS THE STANDARD
C     VALUES FOR EACH OF THE THREE SYSTEMS, 1. DISPERSED PHASE,
C     2. FLUID BED, 3. FIXED BED
C     THESE VALUES WILL BE USED IN THE PROGRAM UNLESS OTHER VALUES ARE
C     REQUESTED
C     VAR 1 = NUMBER OF MEN PER SHIFT
      VAR(1,1) =4.0
      VAR(1,2) = 5.0
      VAR(1,3) = 5.0
C     VAR 2 = SORBENT COST PER POUND , DOLLARS
      VAR(2,1) = 0.25
      VAR(2,2) = 0.25
      VAR(2,3) = 0.25
C     VAR 3 =  COOLING WATER COST, DOLLARS PER GALLON
      VAR(3,1) = 0.00005
      VAR(3,2) = 0.00005
      VAR(3,3) = 0.00005
C     VAR 4 = NATURAL GAS COST , DOLLARS PER MM BTU
      VAR(4,1) = 0.27
      VAR(4,2) = 0.27
      VAR(4,3) = 0.27
C     VAR 5 = POWER COST, DOLLARS PER KILOWATT HOUR
      VAR(5,1) = 0.006
      VAR(5,2) = 0.006
      VAR(5,3) = 0.006
C     VAR 6 = HEAT CREDIT, DOLLARS PER MM BTU
      VAR(6,1) = 0.25
      VAR(6,2) = 0.25
      VAR(6,3) = 0.25
C    VAR 7 = STEAM CREDIT , DOLLARS PER POUND OF STEAM
      VAR(7,1) = 0.00025
      VAR(7,2) = 0.00025
      VAR(7,3) = 0.00025
C     VAR 8 = CAPITAL CHARGES, PERCENTAGE OF FIXED CAPITAL INVESTMENT
      VAR(8,1) = 13.0
      VAR(8,2) = 13.0
      VAR(8,3) = 13.0
C     VAR 9 = FIXED CAPITAL INVESTMENT, LESS SORBENT INVENTORYY COST,
C     DOLLARS
      VAR(9,1) = 30170000.
      VAR(9,2) = 35502000.
      VAR(9,3) = 39725000.
C     VAR 10 = STREAM FACTOR, HOURS PER YEAR
      VAR(10,1) = 5250.
      VAR(10,2) = 5250.
      VAR(10,3) = 5250.
C     VAR 11 = SORBENT INVENTORY,  POUNDS
      VAR(11,1) = 2200000.
      VAR(11,2) = 1592000.
      VAR(11,3) = 1940000.
C     VAR 12 = SORBENT CIRCULATION RATE, POUNDS PER HOUR
      VAR(12,1) = 705792.
      VAR(12,2) = 379800.
C     FOR THE FIXED BED CASE THERE IS NO SORBENT CIRCULATION
C     VAR 13 = SULFUR PRODUCTION, POUNDS PER HOUR
      VAR(13,1) = 17855.
      VAR(13,2) = 17855.
      VAR(13,3) = 19189.
C     VAR 14 = COOLING WATER FLOW, GALLONS PER HOUR
      VAR(14,1) = 301200.
      VAR(14,2) = 301200.
      VAR(14,3) = 526400.
C     VAR 15 = POWER REQUIREMENT, BHP
      VAR(15,1) = 16413.
      VAR(15,2) = 29219.
      VAR(15,3) = 29731.
C     VAR 16 = STEAM PRODUCTION, POUNDS PER HOUR
      VAR(16,1) = 77360.
      VAR(16,2) = 77360.
      VAR(16,3) = 84140.
C     VAR 17 = SULFUR PRICE, DOLLARS PER TON
      VAR(17,1) = 20.0
      VAR(17,2) = 20.0
      VAR(17,3) = 20.0
C     VAR 18 = ATTRITION RATE, PERCENT OF SORBENT CIRCULATION RATE
      VAR(18,1) = 0.1
      VAR(18,2) = 0.05
C     FOR THE FIXED BED CASE THE ATTRITION LOSS IS GIVEN IN TERMS OF
C     YEARS PER INVENTORY REPLACEMENT
      VAR(18,3) = 3.048
C     VAR 19 = COAL CONSUMPTION, TONS PER HOUR
      VAR(19,1) = 330.76895
      VAR(19,2) = 330.76895
      VAR(19,3) = 330.76895
C     VAR 20 = HEAT CREDIT, MM BTU PER HOUR
      VAR(20,1) = 223.232
      VAR(20,2) = 153.114
      VAR(20,3) = 129.17
C     VAR 21 = NATURAL GAS CONSUMPTION, MM BTU PER HOUR
```

```
      VAR(21,1) = 649.7960
      VAR(21,2) = 570.5195
      VAR(21,3) = 618.2887
      INDEX = 0
   37 CALL DATSW(1,NU)
      IF(NU - 1 )30,30,31
   30 KEY = 1
      GO TO 36
   31 CALL DATSW(2,NU)
      IF( NU - 1 )32,32,33
   32 KEY = 2
      GO TO 36
   33 CALL DATSW(3,NU)
      IF( NU - 1 )34,34,35
   34 KEY = 3
      GO TO 36
   35 WRITE(1,200)
  200 FORMAT(2X,37HFOR DISPERSED PHASE PLACE SWITCH 1 UP,/,2X,
     113HFOR FLUID BED,7X,17HPLACE SWITCH 2 UP,/,2X,13HFOR FIXED BED,
     27X,17HPLACE SWITCH 3 UP)
      WRITE(1,201)
  201 FORMAT(1X,/,2X,19HPRESS PROGRAM START)
      PAUSE
      GO TO 37
   36 CONTINUE
      DO 50 J = 1,21
   50 VAR(J,1) = VAR(J,KEY)
      GO TO 8
   61 DO 60 I = 1,21
   60 M(I) = N(I)
      DO 2 I = 1,8
      IF(M(I))2,2,3
    3 K = M(I)
      READ(IR,101)(VAR(I,J),J=1,K)
  101 FORMAT(8F10.6)
    2 CONTINUE
      DO 4 I = 9,16
      IF(M(I))4,4,5
    5 K = M(I)
      READ(IR,102)(VAR(I,J), J = 1,K)
  102 FORMAT(8F10.0)
    4 CONTINUE
      DO 6 I = 17,21
      IF(M(I))6,6,7
    7 K = M(I)
      READ(IR,103)(VAR(I,J),J = 1,K)
  103 FORMAT(5F10.5)
    6 CONTINUE
      DO 8 K = 1,21
      IF(M(K))9,9,8
    9 M(K) = 1
    8 CONTINUE
      J1 = M(1)
      DO 10 K1 = 1,J1
      J2 = M(2)
      DO 10 K2 = 1,J2
      J3 = M(3)
      DO 10 K3 = 1,J3
      J4 = M(4)
      DO 10 K4 = 1,J4
      J5 = M(5)
      DO 10 K5 = 1,J5
      J6 = M(6)
      DO 10 K6 = 1,J6
      J7 = M(7)
      DO 10 K7 = 1,J7
      J8 = M(8)
      DO 10 K8 = 1,J8
      J9 = M(9)
      DO 10 K9 = 1,J9
      J10  = M(10)
      DO 10 K10= 1,J10
      J11  = M(11)
      DO 10 K11= 1,J11
      J12  = M(12)
      DO 10 K12= 1,J12
      J13  = M(13)
      DO 10 K13= 1,J13
      J14  = M(14)
      DO 10 K14= 1,J14
      J15  = M(15)
      DO 10 K15= 1,J15
      J16  = M(16)
      DO 10 K16= 1,J16
      J17  = M(17)
      DO 10 K17= 1,J17
      J18  = M(18)
      DO 10 K18= 1,J18
      J19  = M(19)
      DO 10 K19= 1,J19
      J20  = M(20)
      DO 10 K20= 1,J20
      J21  = M(21)
      DO 10 K21= 1,J21
      IF(INDEX)174,17,174
  174 CALL DATSW(4,LOCK)
      IF(LOCK - 1 )17,17,14
   17 IF( KEY - 1 )37,38,39
   38 WRITE(IW,106)
  106 FORMAT(52X,15HDISPERSED PHASE/)
   39 IF( KEY - 2 )22,40,42
   40 WRITE(IW,107)
  107 FORMAT(55X,9HFLUID BED/)
   42 IF( KEY - 3 )22,43,37
   43 WRITE(IW,199)
  199 FORMAT(55X,9HFIXED BED/)
   22 IF(INDEX)26,26,21
   26 WRITE(IW,141)
```

```
141 FORMAT(53X,13HSTANDARD CASE/)
 21 WRITE(IW,108)
108 FORMAT(42X,36HPROCESS ECONOMICS  -  COMPLETE PLANT///)
 14 FCI = VAR(9,K9) + VAR(11,K11) * VAR(2,K2)
    A1 = VAR(1,K1)*30660.
    A2 = A1 * 0.15
    A3 = FCI * 0.03
    A4 = FCI * 0.02
    A5 = ( A3 + A4 ) * 0.15
    A6A = VAR(3,K3) * VAR(14,K14) * VAR(10,K10)
    A6A1 = VAR(14,K14) / 60.
    A6A2 = VAR(3,K3) * 100000.
    A6B = VAR(4,K4) * VAR(21,K21) * VAR(10,K10)
    A6B1 = VAR(4,K4) * 100.
    A6C = VAR(5,K5) * VAR(15,K15) * VAR(10,K10) * 0.7457
    A6C1 = VAR(15,K15) * 0.7457
    A6C2 = VAR(5,K5) * 100.
    A7A = VAR(6,K6) * VAR(20,K20) * VAR(10,K10)
    A7A1 = VAR(6,K6) * 100.
    A7B = VAR(7,K7) * VAR(16,K16) * VAR(10,K10)
    A7B1 = VAR(7,K7) * 100000.
    A8 = A1 + A2 + A3 + A4 + A5 + A6A + A6B + A6C - A7A - A7B
    A9 = ( A1 + A2 ) * 0.2
    A10 = ( A1 + A2 + A3 + A4 + A5 ) * 0.5
    A11 = A9 + A10
    A12 = FCI * 0.01 * VAR(8,K8)
    IF( KEY - 3 )66,67,66
 67 A13 = VAR(11,K11) * VAR(2,K2) / VAR(18,K18)
    GO TO 68
 66 A13 = VAR(18,K18) * 0.01 * VAR(12,K12) * VAR(10,K10) * VAR(2,K2)
 68 A14 = VAR(17,K17) * VAR(13,K13) * VAR(10,K10) / 2000.
    A15 = A8 + A11 + A13 + A12 - A14
    A16 = A15 / ( VAR(19,K19) * VAR(10,K10) )
    A17 = A15 / (1000. * VAR(10,K10) )
    A18 = A8 + A13 - A14
    IF(INDEX)180,62,180
180 IF(LOCK - 1 )62,62,63
 63 CALL ECHO1
    GO TO 800
 62 CALL ECHO2
800 CONTINUE
    CALL DATSW(0,NU)
    IF(NU - 1 )10,11,10
 10 CONTINUE
    IF(INDEX)13,13,11
 13 INDEX = 1
    GO TO 61
 11 CONTINUE
    CALL EXIT
    END
```

```
// FOR
*ONE WORD INTEGERS
*LIST ALL
* ARITHMETIC TRACE
* TRANSFER TRACE
      SUBROUTINE ECHO1
      DIMENSION VAR(21,10),M(21),N(21)
      COMMON VAR, M, N, FCI, A1,A2,A3,A4,A5,A6A,A6A1,A6A2,A6B,A6B1,A6C
      COMMON A6C1,A6C2,A7A,A7A1,A7B,A7B1,A8,A9,A10,A11,A12,A13,A14,A15
      COMMON A16,A17,A18,K1,K2,K3,K4,K5,K6,K7,K8,K9,K10,K11,K12,K13,K14
      COMMON K15,K16,K17,K18,K19,K20,K21,IR,IW,INDEX,KEY,LOCK
   63 WRITE(IW,170)
  170 FORMAT(45X,29HOPERATING COST FOR THE SYSTEM,/,24X,
     171HIF THE FOLLOWING VALUES ARE SUBSTITUTED FOR THE VALUES IN THE B
     2ASE CASE/)
      DO 64 I = 1,21
      IF(N(I))64,64,65
   65 GO TO (71,72,73,74,75,76,77,78,79,80,81,82,83,84,85,86,87,88,89,
     190,91),I
   71 WRITE(IW,171)I,VAR(I,K1)
  171 FORMAT(10X,5HVAR. ,I2,3X,23HNUMBER OF MEN PER SHIFT,59X,F10.2)
      GO TO 64
   72 WRITE(IW,172)I,VAR(I,K2)
  172 FORMAT(10X,5HVAR. ,I2,3X,31HSORBENT COST PER POUND, DOLLARS,52X,
     1F10.3)
      GO TO 64
   73 WRITE(IW,173)I,VAR(I,K3)
  173 FORMAT(10X,5HVAR. ,I2,3X,38HCOOLING WATER COST, DOLLARS PER GALLON
     1,47X,F10.5)
      GO TO 64
   74 WRITE(IW,174)I,VAR(I,K4)
  174 FORMAT(10X,5HVAR. ,I2,3X,36HNATURAL GAS COST, DOLLARS PER MM BTU,
     147X,F10.3)
      GO TO 64
   75 WRITE(IW,175)I,VAR(I,K5)
  175 FORMAT(10X,5HVAR. ,I2,3X,38HPOWER COST, DOLLARS PER KILO-WATT HOUR
     1,46X,F10.4)
      GO TO 64
   76 WRITE(IW,176)I,VAR(I,K6)
  176 FORMAT(10X,5HVAR. ,I2,3X,31HHEAT CREDIT, DOLLARS PER MM BTU,52X,
     1F10.3)
      GO TO 64
   77 WRITE(IW,177)I,VAR(I,K7)
  177 FORMAT(10X,5HVAR. ,I2,3X,40HSTEAM CREDIT, DOLLARS PER POUND OF STE
     1AM,45X,F10.5)
      GO TO 64
   78 WRITE(IW,178)I,VAR(I,K8)
  178 FORMAT(10X,5HVAR. ,I2,3X,52HCAPITAL CHARGES, PERCENT OF FIXED CAPI
     1TAL INVESTMENT,30X,F10.2)
      GO TO 64
   79 WRITE(IW,179)I,VAR(I,K9)
  179 FORMAT(10X,5HVAR. ,I2,3X,62HFIXED CAPITAL INVESTMENT, LESS SORBENT
     1 INVENTORY COST, DOLLARS,18X,F10.0)
      GO TO 64
   80 WRITE(IW,180)I,VAR(I,K10)
  180 FORMAT(10X,5HVAR. ,I2,3X,29HSTREAM FACTOR, HOURS PER YEAR,51X,
    1F10.0)
      GO TO 64
 81   WRITE(IW,181)I,VAR(I,K11)
181   FORMAT(10X,5HVAR. ,I2,3X,25HSORBENT INVENTORY, POUNDS,55X,F10.0)
      GO TO 64
 82   WRITE(IW,182)I,VAR(I,K12)
182   FORMAT(10X,5HVAR. ,I2,3X,41HSORBENT CIRCULATION RATE, POUNDS PER H
    1OUR,39X,F10.0)
      GO TO 64
 83   WRITE(IW,183)I,VAR(I,K13)
183   FORMAT(10X,5HVAR. ,I2,3X,34HSULFUR PRODUCTION, POUNDS PER HOUR,
    146X,F10.0)
      GO TO 64
 84   WRITE(IW,184)I,VAR(I,K14)
184   FORMAT(10X,5HVAR. ,I2,3X,36HCOOLING WATER FLOW, GALLONS PER HOUR,
    144X,F10.0)
      GO TO 64
 85   WRITE(IW,185)I,VAR(I,K15)
185   FORMAT(10X,5HVAR. ,I2,3X,22HPOWER REQUIREMENT, BHP,58X,F10.0)
      GO TO 64
 86   WRITE(IW,186)I,VAR(I,K16)
186   FORMAT(10X,5HVAR. ,I2,3X,33HSTEAM PRODUCTION, POUNDS PER HOUR,
    147X,F10.0)
      GO TO 64
 87   WRITE(IW,187)I,VAR(I,K17)
187   FORMAT(10X,5HVAR. ,I2,3X,29HSULFUR PRICE, DOLLARS PER TON,53X,
    1F10.2)
      GO TO 64
 88   IF ( KEY - 3 )220,221,220
220   WRITE(IW,188)I,VAR(I,K18)
188   FORMAT(10X,5HVAR. ,I2,3X,51HATTRITION RATE, PERCENT OF SORBENT CIR
    1CULATION RATE,31X,F10.2)
      GO TO 64
221   WRITE(IW,388)I,VAR(I,K18)
388   FORMAT(10X,5HVAR. ,I2,3X,38HATTRITION LOSS, YEARS / INVENTORY LOSS
    1,45X,F10.3)
      GO TO 64
 89   WRITE(IW,189)I,VAR(I,K19)
189   FORMAT(10X,5HVAR. ,I2,3X,31HCOAL CONSUMPTION, TONS PER HOUR,54X,
    1F10.5)
      GO TO 64
 90   WRITE(IW,190)I,VAR(I,K20)
190   FORMAT(10X,5HVAR. ,I2,3X,28HHEAT CREDIT, MM BTU PER HOUR,57X,
    1F10.5)
      GO TO 64
 91   WRITE(IW,191)I,VAR(I,K21)
191   FORMAT(10X,5HVAR. ,I2,3X,40HNATURAL GAS CONSUMPTION, MM BTU PER HO
    1UR,45X,F10.5)
 64   CONTINUE
      WRITE(IW,192)A15
192   FORMAT(//,36X,32HOPERATING COST, DOLLARS PER YEAR,5X,1H$,F10.0/)
      WRITE(IW,193)A16
193   FORMAT(36X,30HDOLLARS PER TON OF COAL BURNED,7X,F10.3)
      WRITE(IW,194)A17
194   FORMAT(36X,32HMILS PER KWH AT 1000 MW CAPACITY,5X,F10.3)
      WRITE(IW,195)
195   FORMAT(1H1)
      RETURN
      END
```

```
// FOR
*ONE WORD INTEGERS
* ARITHMETIC TRACE
* TRANSFER TRACE
* LIST ALL
      SUBROUTINE ECHO2
      DIMENSION VAR(21,10),M(21),N(21)
      COMMON VAR, M, N, FCI, A1,A2,A3,A4,A5,A6A,A6A1,A6A2,A6B,A6B1,A6C
      COMMON A6C1,A6C2,A7A,A7A1,A7B,A7B1,A8,A9,A10,A11,A12,A13,A14,A15
      COMMON A16,A17,A18,K1,K2,K3,K4,K5,K6,K7,K8,K9,K10,K11,K12,K13,K14
      COMMON K15,K16,K17,K18,K19,K20,K21,IR,IW,INDEX,KEY,LOCK
   62 WRITE(IW,109)FCI
  109 FORMAT(30X,33HFIXED CAPITAL INVESTMENT  ( FCI ),5X,1H$,F10.0)
      WRITE(IW,110)VAR(10,K10)
  110 FORMAT(30X,11HSTREAM TIME,5X,14HHOURS PER YEAR,9X,F10.0///)
      WRITE(IW,111)
  111 FORMAT(5X,11HDIRECT COST,86X,8H$ / YEAR/)
      WRITE(IW,112)VAR(1,K1),A1
  112 FORMAT(8X,2H1.,2X,18HOPERATING LABOR ( ,F4.2,28H MEN / SHIFT AT $3
     1.50 / HR ) ,37X,1H$,F10.0)
      WRITE(IW,113)A2
  113 FORMAT(8X,2H2.,2X,34HSUPERVISION - 15 PERCENT OF ITEM 1 ,54X,F10.0
     1)
      WRITE(IW,114)A3
  114 FORMAT(8X,2H3.,2X,36HMAINTENANCE LABOR - 3 PERCENT OF FCI,52X,F10.
     10)
      WRITE(IW,115)A4
  115 FORMAT(8X,2H4.,2X,39HMAINTENANCE MATERIAL - 2 PERCENT OF FCI,49X,
     1F10.0)
      WRITE(IW,116)A5
  116 FORMAT(8X,2H5.,2X,40HSUPPLIES - 15 PERCENT OF ITEMS ( 3 + 4 ),
     148X,F10.0)
      WRITE(IW,117)
  117 FORMAT(8X,2H6.,2X,11HUTILITIES ,)
      WRITE(IW,118)A6A1,A6A2,A6A
  118 FORMAT(15X,2HA.,2X,15HCOOLING WATER -,F6.0,8H GPM AT ,F5.1,
     115H CENTS / M GAL.,32X,F10.0)
      WRITE(IW,119)VAR(21,K21),A6B1,A6B
  119 FORMAT(15X,2HB.,2X,13HNATURAL GAS -,F8.3,18H MM BTU / HOUR AT ,
     1F5.1,15H CENTS / MM BTU,   22X,F10.0)
      WRITE(IW,120)A6C1,A6C2,A6C
  120 FORMAT(15X,2HC.,2X,7HPOWER -,F7.0,7H KW AT ,F4.2,12H CENTS / KWH,
     144X,F10.0)
      WRITE(IW,121)
  121 FORMAT(8X,2H7.,2X,9HCREDITS ,)
      WRITE(IW,122)VAR(20,K20),A7A1,A7A
  122 FORMAT(15X,2HA.,2X,6HHEAT -,F8.3,17H MM BTU / HOUR AT ,F5.1,
     115H CENTS / MM BTU,29X,1H(,F10.0,1H))
      WRITE(IW,123)VAR(16,K16),A7B1,A7B
  123 FORMAT(15X,2HB.,2X,7HSTEAM -,F8.0,18H POUNDS / HOUR AT ,F5.1,
     115H CENTS / M LBS.,27X,1H(,F10.0,1H)/)
      WRITE(IW,124)A8
  124 FORMAT(8X,2H8.,2X,20HSUBTOTAL DIRECT COST,3X,61H - - - - - - - - -
     1 - - - - - - - - - - - - - - - - - - - - ,3X,1H$,F10.0/)
      WRITE(IW,125)VAR(13,K13),VAR(17,K17),A14
  125 FORMAT(8X,2H9.,2X,15HSULFUR CREDIT -,F8.0,18H POUNDS / HOUR AT ,
     1F6.2,8H $ / TON,32X,1H(,F10.0,1H))
      IF( KEY - 3 )23,20,23
   23 WRITE(IW,126)VAR(18,K18),VAR(12,K12),A13,VAR(2,K2)
  126 FORMAT(7X,3H10.,2X,16HATTRITION LOSS -,F6.3,16H PERCENT OF THE ,
     1F9.0,14H POUNDS / HOUR,27X,F10.0,/,35X,20HCIRCULATION RATE AT ,
     2F4.2,10H $ / POUND)
      GO TO 25
   20 WRITE(IW,142)VAR(18,K18),VAR(2,K2),A13,VAR(11,K11)
  142 FORMAT(7X,3H10.,2X,16HATTRITION LOSS -,F6.3,27H YEARS / INVENTORY
     1LOSS AT ,F4.2,10H $ / POUND,25X,F10.0,/,35X,6HAND A ,F10.0,
     216H POUND INVENTORY)
   25 WRITE(IW,128)A18
  128 FORMAT(7X,3H11.,2X,17HTOTAL DIRECT COST,3X,65H - - - - - - - - - -
     1 - - - - - - - - - - - - - - - - - - - - - - ,2X,1H$,F10.0/)
      WRITE(IW,129)
  129 FORMAT(5X,13HINDIRECT COST/)
      WRITE(IW,130)A9
  130 FORMAT(7X,3H12.,2X,42HPAYROLL OVERHEAD - 20 PERCENT OF ( 1 + 2 ),
     146X,F10.0)
      WRITE(IW,131)A10
  131 FORMAT(7X,3H13.,2X,52HPLANT OVERHEAD - 50 PERCENT OF ( 1 + 2 + 3 +
     1 4 + 5 ),36X,F10.0)
      WRITE(IW,132)A11
  132 FORMAT(7X,3H14.,2X,19HTOTAL INDIRECT COST,3X,62H - - - - - - - - -
     1 - - - - - - - - - - - - - - - - - - - - - -,3X,1H$,F10.0/)
      WRITE(IW,133)
  133 FORMAT(5X,10HFIXED COST/)
      WRITE(IW,134)VAR(8,K8),A12
  134 FORMAT(7X,3H15.,2X,17HCAPITAL CHARGES -,F6.2,16H PERCENT OF FCI,
     148X,1H$,F10.0)
      WRITE(IW,135)
  135 FORMAT(15X,66H( INCLUDES DEPRECIATION, INSURANCE, TAXES, COST OF C
     1APITAL, ETC. )//)
      WRITE(IW,136)
  136 FORMAT(5X,20HTOTAL OPERATING COST/)
      WRITE(IW,137)A15
  137 FORMAT(7X,3H16.,2X,45HTOTAL OPERATING COST - ITEMS ( 11 + 14 + 15
     1),39H - - - - - - - - - - - - - - - - - - - ,3X,1H$,F10.0/)
      WRITE(IW,138)
  138 FORMAT(5X,9HUNIT COST/)
      WRITE(IW,139)VAR(19,K19),A16
  139 FORMAT(7X,3H17.,2X,16H$ / TON COAL AT ,F8.3,12H TONS / HOUR,53X,
     1F8.3)
      WRITE(IW,140)A17
  140 FORMAT(7X,3H18.,2X,30HMILS / KWH AT 1000 MW CAPACITY,59X,F8.3)
      WRITE(IW,198)
  198 FORMAT(1H1)
      RETURN
      END
```

IX. APPENDIX

For reference purposes, the base case process description and flowsheets are reproduced here. These items are taken from the phases 1 and 2 final report (February 28, 1970) which should be referred to for any additional details that may be required. The following items are included in this appendix:

1. Dispersed Phase
2. Fluid Bed
3. Fixed Bed
4. Regeneration Plant
5. Claus Plant Pretreatment

It should be noted that the figure and table numbers in this section are the ones used in the phases 1 and 2 report and therefore are not in sequence with those used in this report.

a. Dispersed Phase

(1) Dispersed Phase Process Description: Flow Sheet P3197-D

The scheme is based on removing 90% of the SO_2 in the flue gases from a 1000 MW power plant burning coal containing 3% sulfur. It should be noted that the quantities shown are for only one train of equipment whereas two parallel trains are required when the power plant is running at full capacity.

Hot flue gases (about 600°F) from the power plant enter the bottom of the dispersed phase reactor, D-1, and flow upward through the reactor at about 25 feet per second with a total gas residence time of about three seconds being achieved. In the reactor the flue gases contact the sorbent particles under conditions designed to effect removal of 90% of the SO_2 present in the entering flue gas. The treated flue gas effluent from the top of the reactor passes through a one-stage cyclone to recover sorbent particles greater than about 60 mesh, and then it sent to the power plant stack after first flowing through the air preheater and electrostatic precipitator. An alternative design would relocate the electrostatic precipitator upstream of the SO_2 removal plant. The air preheater and electrostatic precipitator are standard pieces of equipment in power plant operations and are not part of the SO_2 removal process.

Sorbent is introduced into the D-1 reactor at three different points: a heavy particle recycle stream near the top of the reactor, a fines recycle stream which enters near the bottom of the reactor, and a stream of regenerated sorbent also near the bottom. The heavy particle stream comprises particles whose settling velocity is greater than the superficial gas velocity in the reactor, and thus these particles fall downward through the reactor and are withdrawn from the bottom. The regenerated solids consist of a range of particle sizes, including particles with settling velocities both greater and less than the reactor gas velocity. The heavier particles settle to the bottom and are withdrawn through the bottom exit line while the lighter particles are carried upward and are recovered in the reactor cyclones. The fines recycle stream is comprised of a portion of the fines recovered in the reactor cyclones.

The bottom exit solids are transported via a dense phase riser line to the top of the reactor where the fluidizing air and solids are separated in the gas-solids separator, L-2. The air leaving L-2 is combined with the flue gas leaving D-1 for eventual discharge to the atmosphere via the power plant stack. The solids in the bottom of L-2 are maintained in a fluidized state and are withdrawn through a standpipe and split into two streams: a recycle stream to D-1, and a side stream which is sent to the regeneration section.

The lighter sorbent particles carried overhead with the flue gas move through the reactor at different velocities depending on the slip velocity of the individual particles (i.e., slip velocity is the difference in gas velocity and particle settling velocity), and therefore have different residence times in the reactor. Thus the heavier of the overhead particles have the lowest slip and hence highest residence time. The lightest particles have the shortest residence time. However, these short residence time particles have a high surface area to weight ratio. Therefore, it has been assumed that most of these overhead particles will have either sufficient residence time or sufficient exposure of surface area such that they will be loaded to design capacity during one pass through the reactor and will not have to be recycled. In case one pass is not sufficient, however, provision for recycle of some of the fines back to the reactor has been included on the flowsheet.

Entrained fines in the flue gas effluent from the top of D-1 are recovered in cyclones L-1A-1F. The large volume of gases to be handled require the use of several cyclones in parallel. The flue gas effluent from these cyclones is sent to the power plant stack, while the recovered fines (+60 mesh particles) may be recycled to the reactor or may be combined with the side stream of heavier particles from L-2 and sent to the regeneration section as discussed above.

The regeneration section comprises a solids heater, D-2 and a regenerator, D-3. These vessels are operated at a nominal pressure of 30 psia (selection of the operating pressure is discussed later), whereas the D-1 reactor pressure is essentially atmospheric. Thus, it is necessary to feed solids from a system at low pressure into one at high pressure and vice versa. The latter condition can be met easily by taking a suitable pressure drop across the solids feeder, but the former requires a build-up of pressure in the solids transfer line between the D-1 reactor and solids heater D-2. This has been done by use of a dense phase solids standpipe extending from the bottom of the elevated L-2 gas-solids separator to grade. The height of the standpipe is such that the pressure at grade is equivalent to that in the solids heater. Air is injected into the line at grade such that the solids are pneumatically transported to the solids heater and fed into the top bed through a feed nozzle designed to separate the air and solids so that bed stability is not affected. The transporting air mixes with the gases leaving the top bed, the combined gas stream then passes through an internal cyclone for fines removal, and finally joins the main flue gas exit stream. If a single bed heater were used, the solids from D-1 could be injected directly into the bed and the transporting air would then supply part of the combustion air thus reducing the air compressor horsepower. A two stage heater proved to be more economical, however, since the reduced fuel requirements more than offset the horsepower savings.

Sorbent solids are heated from 658°F to 1400°F in solids heater D-2, which contains two fluidized beds: (1) a top bed which preheats the cold solids by exchange with hot exit gases from the bottom bed, thereby effecting heat economy, and (2) a bottom fluidized bed in which the solids are heated to the desired temperature of 1400°F by direct combustion of natural gas in the bed itself. Hot solids leave the bottom bed through an overflow withdrawal line and are fed into the moving bed regenerator, D-3. A slide valve in the discharge line from D-2 and a splash baffle underneath the end of the line in D-3 provide a seal to prevent the spent regeneration gas from flowing into the D-2 solids heater.

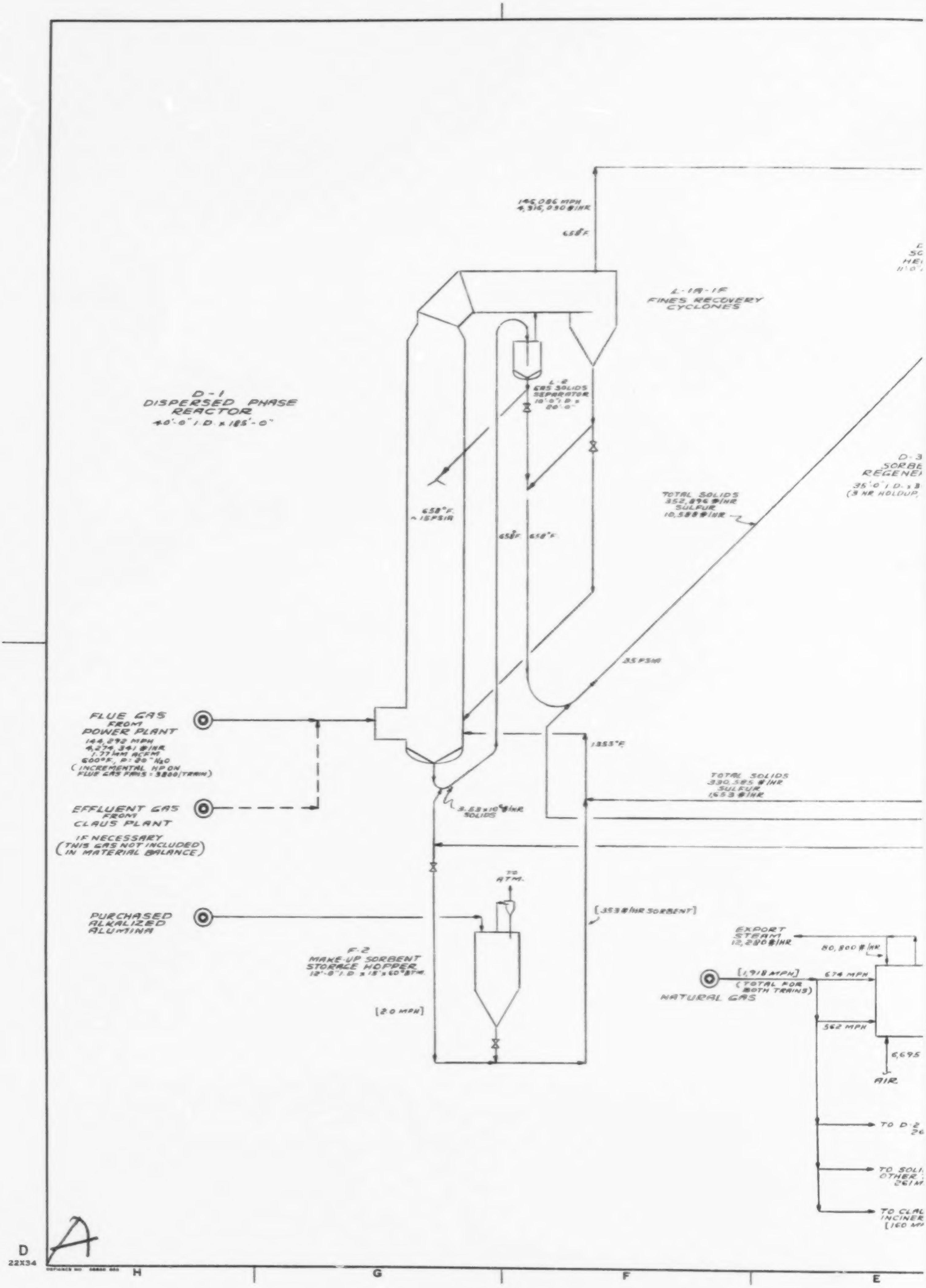
Sorbent is regenerated in D-3 by countercurrent contact with upflowing regeneration gas. The regeneration gas is supplied to the regenerator from the steam-methane reformer, L-3. The reformer gas generation process flowsheet (P1374-B) and description are given in Section B.2.e. part (1) of this report.

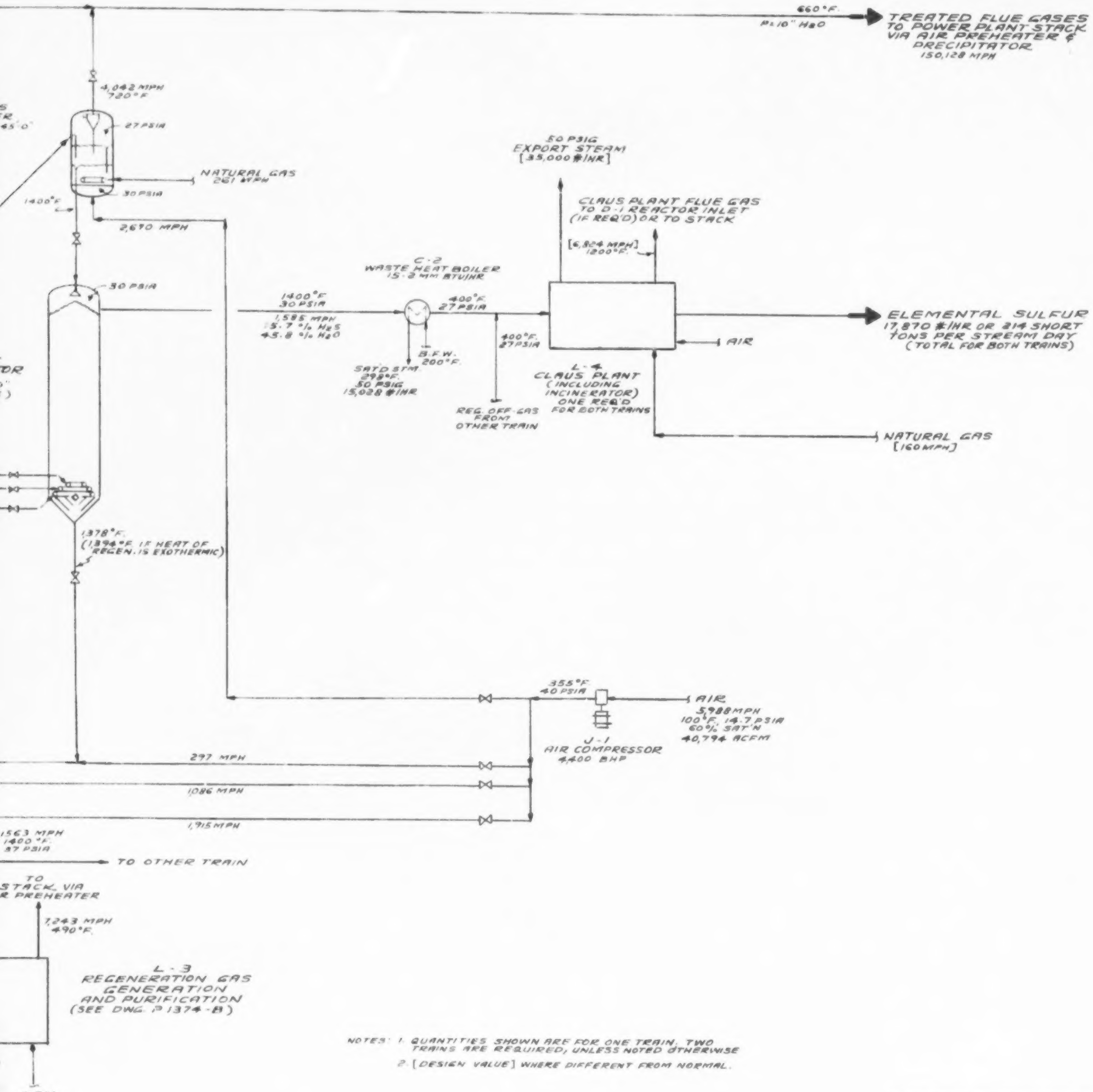
The sorbent which is loaded with sulfur in the form of SO_4 or SO_3 is converted back to its original state ($NaAlO_2$) by the reducing atmosphere in the regenerator. The 1400°F regenerator off-gas which contains the reduced sulfur in the form of H_2S exits from the top of the regenerator and after passing through the waste heat boiler, C-2, it is routed to the Claus Plant for the recovery of elemental sulfur. A description of the Claus Plant is given in section 6, part (3) of this report.

The regenerator is designed to provide (1) a bed which moves down through the vessel in a plug-type flow pattern, and (2) uniform gas distribution throughout the bed. These objectives are achieved by (1) installing suitable baffles near the vessel bottom, and (2) providing gas distribution rings spaced so that equal bed heights are obtained above each ring by allowing for build-up of sorbent due to a single inlet line and the angle of repose of the sorbent.

Hot, regenerated sorbent is withdrawn through a standpipe and is pneumatically transported to the bottom of reactor D-1, thus completing the solids cycle. Make-up sorbent is stored in hopper F-2, and added to the system as required by feeding it in through the regenerated solids transfer line.

The air required for the solids heater D-2 and that needed for fluidizing and transporting purposes, is supplied by the centrifugal air compressor J-1. The J-1 discharge stream is split into four separate streams: the three streams used for transporting the solids and the stream to the solids heater.


146,086 MPH
4,316,030 #/HR
658°F
L-1A-1F
FINES RECOVERY
CYCLONES
D-1
DISPERSED PHASE
REACTOR
40'-0" I.D. x 185'-0"
L-2
GAS SOLIDS
SEPARATOR
10'-0" I.D. x
20'-0"
658°F
~15PSIA
658°F
658°F
TOTAL SOLIDS
352,896 #/HR
SULFUR
10,588 #/HR
D-3
SORBE
REGENE
35'-0" I.D. x 3
(3 HR HOLDUP,
D
SC
HE
11'-0"
35 PSIA
FLUE GAS
FROM
POWER PLANT
144,292 MPH
4,274,341 #/HR
1.77 MM ACFM
600°F, P: 20" H2O
(INCREMENTAL HP ON
FLUE GAS FANS: 3800/TRAIN)
1355°F
EFFLUENT GAS
FROM
CLAUS PLANT
IF NECESSARY
(THIS GAS NOT INCLUDED
IN MATERIAL BALANCE)
3.53 x 10⁶ #/HR
SOLIDS
TOTAL SOLIDS
330,585 #/HR
SULFUR
1,653 #/HR
TO
ATM.
PURCHASED
ALKALIZED
ALUMINA
[353 #/HR SORBENT]
F-2
MAKE-UP SORBENT
STORAGE HOPPER
12'-0" I.D. x 15' x 60° BTM.
[2.0 MPH]
EXPORT
STEAM
12,280 #/HR
80,800 #/HR
[1,918 MPH]
(TOTAL FOR
BOTH TRAINS)
674 MPH
NATURAL GAS
562 MPH
6,695
AIR
TO D-2
26
TO SOLI
OTHER
261 M
TO CLAU
INCINER
[160 M
A
D
22X34
H
G
F
E


660°F.
P=10" H₂O
TREATED FLUE GASES TO POWER PLANT STACK VIA AIR PREHEATER & PRECIPITATOR
150,128 MPH
4,042 MPH
720°F
27 PSIA
NATURAL GAS
261 MPH
30 PSIA
1400°F
2,670 MPH
30 PSIA
50 PSIG
EXPORT STEAM
[35,000 #/HR]
CLAUS PLANT FLUE GAS TO D-1 REACTOR INLET (IF REQ'D) OR TO STACK
[6,824 MPH]
1200°F.
C-2
WASTE HEAT BOILER
15.2 MM BTU/HR
1400°F
30 PSIA
1,585 MPH
15.7 % H₂S
45.8 % H₂O
400°F
27 PSIA
400°F.
27 PSIA
B.F.W.
200°F.
SAT'D STM.
298°F.
50 PSIG
15,028 #/HR
L-4
CLAUS PLANT
(INCLUDING INCINERATOR)
ONE REQ'D FOR BOTH TRAINS
REG. OFF-GAS FROM OTHER TRAIN
AIR
ELEMENTAL SULFUR
17,870 #/HR OR 214 SHORT TONS PER STREAM DAY
(TOTAL FOR BOTH TRAINS)
NATURAL GAS
[160 MPH]
1378°F.
(1394°F IF HEAT OF REGEN. IS EXOTHERMIC)
355°F
40 PSIA
AIR
5,988 MPH
100°F. 14.7 PSIA
60% SAT'N
40,794 ACFM
J-1
AIR COMPRESSOR
4,400 BHP
297 MPH
1086 MPH
1,915 MPH
1563 MPH
1400°F.
37 PSIA
TO OTHER TRAIN
TO STACK VIA
R PREHEATER
7,243 MPH
490°F.
L-3
REGENERATION GAS GENERATION AND PURIFICATION
(SEE DWG. P 1374-B)
B.F.W.
DS HEATER
EATER
ANT

NOTES: 1. QUANTITIES SHOWN ARE FOR ONE TRAIN. TWO TRAINS ARE REQUIRED, UNLESS NOTED OTHERWISE

2. [DESIGN VALUE] WHERE DIFFERENT FROM NORMAL.

NO.	DESCRIPTION	DATE	BY	APPROVED
1	GENERAL	2-27-70	A.G.S.	

REVISIONS

ENGINEERED A.G.S.
DRAWN F.R.
RELEASED FOR DESIGN
DATED 8-31-69
APPROVED
ISSUED FOR CONSTRUCTION

THE M. W. KELLOGG COMPANY
a division of PULLMAN INCORPORATED

ALKALIZED ALUMINA PROCESS
DISPERSED PHASE FLOW SHEET
FOR
1,000 MW POWER PLANT FLUE GASES

CLASS & ITEM	AREA	JOB NUMBER	DRAWING NUMBER
		6045	P3197-D

DRAWING NO.

B

b. Fluid Bed

(1) Fluid Bed Process Description: Flow Sheet P3198-D

The following process description is for the alkalized alumina process scheme shown on flow sheet P3198-D. The scheme is based on removing 90% of the SO_2 in the flue gases from a 1000 MW power plant burning a 3% sulfur coal.

The equipment shown on this flow sheet is designed to handle 50% of the power plant effluent. The remaining 50% of the power plant effluent is fed to a second train of equipment identical to that shown on this flow sheet. In actual plant construction, however, the gas flow will not be split until after the flue gas booster fans. This will enable the use of a single inlet header, thereby decreasing duct construction costs. The same practice will be used for the sorber effluent gases with a single duct being provided at this point for the same reason as stated above. The flow sheet, therefore, schematically represents only the flows in one train of equipment.

The SO_2-rich power plant effluent enters the flow sheet at 600°F and 10 in. H_2O after first passing through the power plant electrostatic precipitator. This gas is then compressed to 45 in. H_2O by the flue gas booster fan, J-1. The compression is required to overcome line loses and sorber pressure drop and enables the gas to be returned to the power plant circuit at the same pressure at which it was withdrawn. The compressed gases flow through the common inlet header and are then split into parallel feeds for each of the five fluid bed sorbers in the train. The gases pass through fluidized beds of alkalized alumina contained within these vessels wherein SO_2 removal is effected.

The fluid bed sorbers, D-1 a thru e, are designed for 90% removal of SO_2. The number and height of these beds were determined by the CEGB-Blyth, and USBM-Albany sorption models, with appropriate safety factors. The SO_2-depleted effluent flue gas is then passed through internal cyclones to minimize sorbent losses. The exit gases from the sorbers enter the common exit gas header and are returned to the power plant circuit where they flow through the power plant air preheater for heat recovery before being discharged to the atmosphere via the stack.

The sulfur loaded sorbent leaves the sorbers via a solids overflow system and is then pneumatically conveyed to an intermediate feed hopper. This feed hopper, the gas-solids separator, L-1, serves three functions: first, it provides a common hopper for the solids discharge from each of the sorbers in the train; second, it provides a six minute holdup based on solids feed to the regeneration train, thus stabilizing possible fluctuations in flow; third, it provides enough elevation for a dense phase dipleg to build up sufficient pressure (30 psia) to match that in the regeneration train.

The sulfur loaded sorbent leaves the gas-solids separator, L-1, through a dense phase fluidized standpipe and flows down 90 feet to a reverse loop where it is joined by a transport air line for pneumatic conveyance (@ 1 lb/cu ft loadings, throughput basis) to the regeneration train. The elevations of the regeneration train's solids heater and the gas-solids separator were set by a pressure balance around the pneumatic transport loops. A solids loading of two pounds per cubic foot was used in this calculation. This agrees with the slip factor of two which was estimated from the particle size distribution and theoretical settling velocities. The transport gas used in both loops is air at 35 psia.

The sulfur loaded solids enter the upper bed of the solids heater, D-3, at approximately 634°F. Two fluidized beds were used in this vessel to allow the 1400°F gaseous effluent from the lower bed to preheat the solids in the upper bed. This arrangement minimizes the natural gas and air requirements for the system by decreasing the duty required in the lower bed. The preheated solids leave the upper bed at 726°F via an overflow standpipe and are fed to the lower bed. The major part of the heating duty is supplied in this lower bed where the solids are heated by the direct combustion of natural gas and air. The solids leave the lower bed at the gas-solids equilibrium temperature, 1400°F, and exit via an overflow standpipe which feeds the sorbent regenerator, D-2.

The effluent gas from the upper bed of the heater mixes with the pneumatic transport air in the vapor space above the upper bed. These gases are then passed through an internal cyclone to separate the entrained solids from the gas stream. The mixed gases leave the vessel at 702°F and are let down by a valve from approximately 25 psia to 10 in. H_2O. The combined heater effluent then flows to the main flue gas exit header for heat recovery and ultimate disposal through the power plant stack.

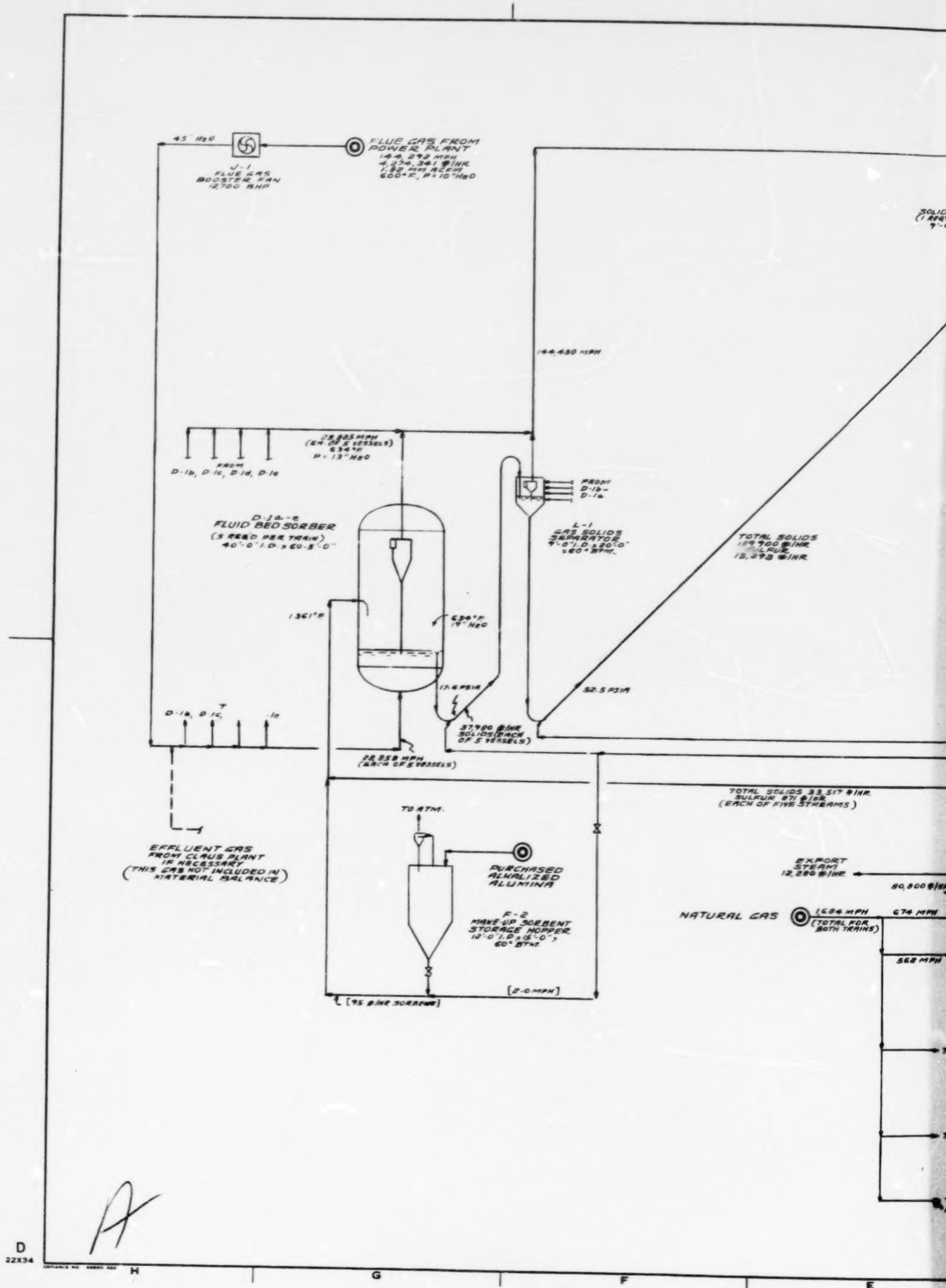
The sorbent leaves the solids heater at 1400°F and 30 psia and enters the fluid bed sorbent regenerator, D-2. In this vessel the sulfated alkalized alumina sorbent is regenerated with a reducing gas to desulfurize the sorbent from 7 weight percent sulfur to 2.6 weight percent. The sulfur-rich reducing gas effluent formed in the regenerator is passed through a cyclone to remove entrained solids and then sent to a Claus unit for sulfur recovery as elemental sulfur. A separate discussion of the Claus unit is given in section B.2.e, Part (3) of this report.

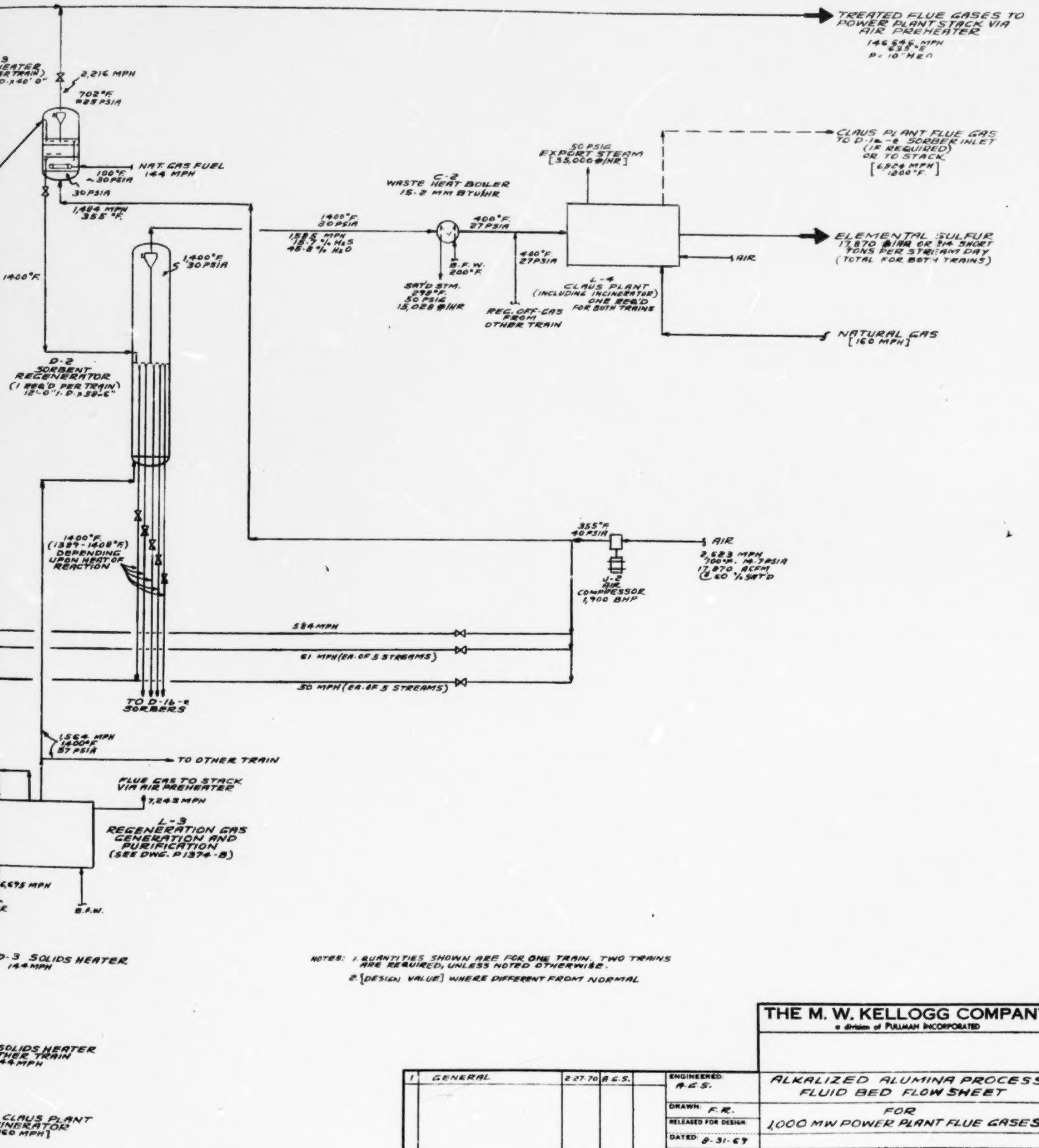
The regeneration gas is supplied to the regenerator from the steam-methane reformer, L-3. The reformer gas generation process flow sheet (P1374-B) and description are given in section B.2.e.,Part (1) of this report. The regeneration gas at 1400°F and 25% excess over stoichiometric proportions enters the bottom of the vessel and passes through the fluidized sorbent bed, removing the sulfur from the sorbent. The products of the reduction reactions, based on a reformed natural gas feed are: sodium aluminate ($NaAlO_2$), hydrogen sulfide (H_2S), carbonyl sulfide (COS), sulfur dioxide (SO_2), elemental sulfur (S_2), and water. These gaseous products of the reactions leave overhead, while the sulfur depleted solids are withdrawn through a fluidized standpipe.

The solids leave the sorbent regenerator, D-2, at an average temperature of about 1400°F via five overflow standpipes. These standpipes are fluidized and are provided with control valves to regulate the flow in each line. The lines serve as the dense phase dipleg for a pneumatic air transport system which carries the sorbent from the standpipes to the fluid bed sorbers, one line being used to feed each sorber. The regenerated sorbent then enters the sorbers and the cycle is completed.

Along with the major process loop of sorption, heating and regeneration, two other systems are shown on the flow sheet. These are the air and sorbent make-up systems. In the air system, air is compressed from atmospheric conditions to 40 psia by the air compressor, J-2. The air flow is then split into four streams: solids transport to the sorbers, solids transport to the gas-solids separator, solids transport to the solids heater and combustion air to the solids heater. The design temperature on the compressor effluent is 355°F. Thus the transport air modifies the temperature of the solids somewhat. For example, the regenerated solids are cooled from 1400°F to 1361°F by the transport air.

A make-up sorbent storage hopper, F-2, is provided to replace solids lost to the flue gas effluent through attrition. The vessel was designed to provide 1000 cubic feet of active storage capacity and thus be able to receive truck load quantities of purchased alkalized alumina sorbent. The solids discharge from the hopper is connected to the pneumatic transport system of a sorber feed line. Flows of make-up sorbent shown on the flow sheet are estimated since attrition data for this type of system are not available.


45" H2O
J-1
FLUE GAS
BOOSTER FAN
12700 BHP
FLUE GAS FROM
POWER PLANT
144,292 MPH
4,274,341 #/HR
1.82 MM ACFM
600°F, P=10"H2O
144,480 MPH
28,825 MPH
(EA. OF 5 VESSELS)
634°F
P=13" H2O
FROM
D-1b, D-1c, D-1d, D-1e
FROM
D-1b-
D-1e
D-1a-e
FLUID BED SORBER
(5 REQ'D PER TRAIN)
40'-0" I.D. x 60-5'-0"
L-1
GAS SOLIDS
SEPARATOR
9'-0" I.D. x 20'-0"
x 60° BTM.
TOTAL SOLIDS
159,700 #/HR
SULFUR
13,293 #/HR
SOLID
(1 REQ'
9'-
1361°F
634°F
19" H2O
17.6 PSIA
32.5 PSIA
D-1b, D-1c, ·1e
37,900 #/HR
SOLIDS (EACH
OF 5 VESSELS)
28,858 MPH
(EACH OF 5 VESSELS)
TOTAL SOLIDS 33,517 #/HR
SULFUR 871 #/HR
(EACH OF FIVE STREAMS)
TO ATM.
EFFLUENT GAS
FROM CLAUS PLANT
IF NECESSARY
(THIS GAS NOT INCLUDED IN
MATERIAL BALANCE)
PURCHASED
ALKALIZED
ALUMINA
F-2
MAKE-UP SORBENT
STORAGE HOPPER
12'-0" I.D. x 15'-0",
60° BTM.
[2.0 MPH]
[95 #/HR SORBENT]
EXPORT
STEAM
12,280 #/HR
80,800 #/H
NATURAL GAS
1,684 MPH
(TOTAL FOR
BOTH TRAINS)
674 MPH
562 MPH
A
D
22X34
H
G
F
E


TREATED FLUE GASES TO POWER PLANT STACK VIA AIR PREHEATER
146,646 MPH
635°F
P= 10" H₂O
3
HEATER
ER TRAIN)
D.×40' 0"
2,216 MPH
702°F
≈25 PSIA
NAT. GAS FUEL
144 MPH
100°F
~30 PSIA
30 PSIA
1,484 MPH
355°F
1400°F
1,400°F
30 PSIA
1400°F
30 PSIA
158.5 MPH
15.7 % H₂S
45.8 % H₂O
C-2
WASTE HEAT BOILER
15.2 MM BTU/HR
400°F
27 PSIA
B.F.W.
200°F
SAT'D STM.
298°F
50 PSIG
15,028 #/HR
400°F
27 PSIA
REG. OFF-GAS FROM OTHER TRAIN
50 PSIG
EXPORT STEAM
[35,000 #/HR]
CLAUS PLANT FLUE GAS TO D-16-e SORBER INLET (IF REQUIRED) OR TO STACK
[6,824 MPH
1200°F]
L-4
CLAUS PLANT
(INCLUDING INCINERATOR)
ONE REQ'D FOR BOTH TRAINS
AIR
ELEMENTAL SULFUR
17,870 #/HR OR 214 SHORT TONS PER STREAM DAY
(TOTAL FOR BOTH TRAINS)
NATURAL GAS
[160 MPH]
D-2
SORBENT REGENERATOR
(1 REQ'D PER TRAIN)
12'-0" I.D.×58'-6"
1400°F
(1389-1408°F)
DEPENDING UPON HEAT OF REACTION
355°F
40 PSIA
AIR
2,623 MPH
100°F, 14.7 PSIA
17,870 ACFM
@ 60 % SAT'D
J-2
AIR COMPRESSOR
1,700 BHP
584 MPH
61 MPH (EA. OF 5 STREAMS)
50 MPH (EA. OF 5 STREAMS)
TO D-16-e SORBERS
1,564 MPH
1400°F
57 PSIA
TO OTHER TRAIN
FLUE GAS TO STACK VIA AIR PREHEATER
7,243 MPH
L-3
REGENERATION GAS GENERATION AND PURIFICATION
(SEE DWG. P1374-B)
6,695 MPH
B.F.W.
D-3 SOLIDS HEATER
144 MPH
SOLIDS HEATER
THER TRAIN
44 MPH
CLAUS PLANT
INERATOR
60 MPH]

NOTES: 1. QUANTITIES SHOWN ARE FOR ONE TRAIN. TWO TRAINS ARE REQUIRED, UNLESS NOTED OTHERWISE.

2. [DESIGN VALUE] WHERE DIFFERENT FROM NORMAL

NO.	DESCRIPTION	DATE	BY	APPROVED
1	GENERAL	2-27-70	A.G.S.	

REVISIONS

THE M. W. KELLOGG COMPANY
a division of PULLMAN INCORPORATED

ENGINEERED: A.G.S.
DRAWN: F.R.
RELEASED FOR DESIGN:
DATED: 8-31-69
APPROVED:
ISSUED FOR CONSTRUCTION

ALKALIZED ALUMINA PROCESS
FLUID BED FLOW SHEET
FOR
1,000 MW POWER PLANT FLUE GASES

CLASS & ITEM:
AREA:
JOB NUMBER: 6045
DRAWING NUMBER: P3198-D

DRAWING NO.

B

c. Fixed Bed

(1) Fixed Bed Process Description: Flow Sheet P3196-D

The SO_2-rich flue gas from a 1000 megawatt power plant burning a 3% sulfur coal enters the flow sheet after leaving the power plant's precipitator at 600°F, as stream 1. Stream 1 is compressed by the flue gas booster fan, J-1, from 10 inches of water to 45 inches of water to overcome the line losses and bed pressure drop in the SO_2 removal scheme, calculated to be 35 inches of water; thus the flue gas is returned to the power plant circuit at the same pressure at which it was withdrawn. The compressed gas then enters a header, common to the four fixed bed sorbers, D-1 thru 4. During steady state operation, two of the four sorbers will serve to remove SO_2 from the flue gas, while the remaining two sorbers will be in the regeneration, heating or cooling cycle. A sorber by-pass line to the power plant stack via the air preheater is also provided on the header. This line is designed to accept 50% of the stream 1 flow should maintenance require one of the sorbers to be shut down temporarily.

The stream 1 flow from the header is split and fed to each of the two reactors in which sorption is occurring (in the case shown on the flow sheet sorbers D1 and D2) as streams 1-A & 1-B. Each sorber is composed of 84 annular sorbent beds of alkalized alumina enclosed in a rectangular chamber. The beds are 3'-7" O.D. by 2'-4" I.D. by 22'-0" high, giving a bed thickness of ~7" and a bed pressure drop of 30 in. H_2O, and are described in the design notes.

The flue gas from stream 1 flows into the rectangular sorber chamber then through the annular sorbent beds where SO_2 is removed from the gas stream. The clean flue gas then flows up the center cylinder, formed by the annulus, to a common sorber exit gas chamber and out of the sorber to the effluent flue gas header, stream 2. The clean flue gases flow through this header and combine with the sorber heating or cooling gas from a third sorber. A side stream, stream 4, is withdrawn from these mixed gases for use as a heating or cooling medium for the sorbent beds. The remainder of these mixed gases, stream 3, leaves the process area and flows to the power plant air preheater for heat recovery before being discharged through the stack.

Because the process utilizes fixed bed sorption and regeneration, the operation is on a batch rather than a continuous basis. The cycle time for this process was set by allowing two hours as the time required for the regeneration step (and one hour each for heating and cooling the beds to and from the regeneration temperature of 1200°F. Thus a total heating-regenerating cooling time of four hours is obtained. Based on the above, plus a four hour sorption time, a complete cycling of the sorbers is made in eight hours. Four sorbers and one hour heating and cooling time were chosen to allow major process equipment (such as the recycle flue gas blower, J-2, the regeneration gas generation and purification sections, and the Claus-gas purification section) to operate on a continuous basis.

As mentioned earlier, the sorbent beds must be preheated from the 600°F sorption temperature to the 1200°F level required for the regeneration reaction. Recycled flue gas from the sorber exit header is used as the heating medium for this purpose, and is taken off as a side stream from the effluent flue gas heater, stream 2. This side stream of recycle flue gas, stream 4, is compressed from 10 in. H_2O to 25 in. H_2O, by the recycle flue gas blower, J-2, to overcome line losses and bed pressure drop, which are smaller than for the flue gas circuit because of the smaller gas flow. The recycle gas at 602°F is then passed through the convection section of a natural gas fired furnace where the furnace's combustion products are mixed with the recycle gas. This furnace, the recycle flue gas heater, B-1, was designed to give an effluent flue gas temperature of 1225°F, when using 10% excess air. Two blowers, the air blower, J-3, and the pilot air blower, J-4, provide the air required for combustion in this furnace. Both blowers are designed to provide a positive pressure of 25 in. H_2O, which is sufficient to move the heating gas through the fixed sorbent beds.

The pilot air blower, J-4, serves the function of providing combustion air, stream 5-A, during the cooling cycle when no process flow is required through the furnace. The combustion products from the pilot burner serve to maintain the operation temperature in the furnace, thus preventing thermal cycling in the furnace during the no-flow condition. The effluent from the pilot burner is fed directly to the stack through its own exhaust duct which ties into the clean flue gas duct, stream 3, downstream of the sorbers.

The recycle heating gas effluent from the furnace, stream 7, enters the common sorber inlet duct, stream 1-D, and flows through the sorbent beds, preheating them to 1200°F for regeneration. Heat transfer in the fixed bed is very rapid with the temperature profile in the bed appearing as a sharp wave-front. Because of this, the exit gas temperature is 600°F, the initial bed temperature, for almost the entire heating cycle. It is only during the end of the cycle that the wave-front begins to break through causing an increase in the exit gas temperature. The average bed exit temperature is 625°F, whereas the final bed exit temperature is approximately 1200°F. The spent heating gas then leaves the sorber via the common sorber effluent duct, stream 2-D, and flows into stream 2 for disposal to the stack.

The cooling cycle is essentially the same as the heating cycle; i.e., clean flue gas from stream 2 is recycled to the sorbers to cool the beds from 1200°F to the sorption temperature. There are, however, two differences in this cycle; first, the recycle flue gas heater, B-1, is by-passed; second, the bed exit temperature for the cooling gas is 1200°F for the greater part of the cycle. Due to the latter difference the average effluent gas temperature is increased which means that the recycled gas cannot cool the beds down to the sorption temperature of 600°F, but rather to the mix temperature of the gas in the effluent header, 675°F. To compensate for this, the gas flow rate is increased slightly thus causing an earlier break-through. Once break-through occurs, the effluent gas temperature drops off sharply, giving a mix temperature in the effluent gas header of approximately 600°F. Therefore, during the remainder of the cycle gas inlet cooling gas to the sorber is at the desired temperature. This cooler gas removes further heat by forming a second wave-front in the bed. This wave front precools the front section of the bed at 600°F for sorption and propagates itself as further 600°F gas is passed through the bed. Even if the entire bed has not been cooled to 600°F at the end of the cooling cycle, the introduction of 600°F flue gas for sorption will continue to cool the sorbent by extending the cooling wave-front ahead of the reaction zone.

The last operation cycle is that of regeneration. The sulfated alkalized alumina is regenerated with a reducing gas to desulfurize the alumina and form a sulfur-rich gas which can be fed to a Claus unit for sulfur recovery as elemental sulfur. On the present flow sheet, a 1200°F reformed natural gas, containing 10% CO, was used as the reducing agent. The regeneration gas generation and purification stages are shown on a separate flow sheet, P1374-B, section B.2.e. part (1) of this report. Similarly, the regeneration gas effluent from the sorbers has to be pre-treated prior to entering the Claus unit; the Claus-gas purification stage is also shown on a separate flow sheet, P1375-B, and discussed in section B.2.e. part (2) of this report.

The assumed regeneration gas, containing a 25% excess over stoichiometric proportions, flows as stream 8 to the common sorber inlet duct, stream 1-C, at 1200°F. Leaving the inlet duct, the gas passes into the sorber and through the annular sorbent beds to remove the sulfur from the sorbent. The products of this reduction reaction, based on a natural gas reformate, are: sodium aluminate ($NaAlO_2$), hydrogen sulfide (H_2S), carbonyl sulfide (COS), sulfur dioxide (SO_2), elemental sulfur (S_2), and water. This sulfur-rich gas flows out of the common sorber exit duct, stream 2-C, to the regeneration gas effluent header, stream 9. This header feeds the Claus gas purification stage where the gas is pre-treated (H_2O removal, etc.) and sent to the Claus unit for sulfur recovery as elemental sulfur.

The high process temperatures encountered in the large ducts require water-cooled closure rings (i.e., valve seats) to minimize thermal stresses in the duct walls. To avoid scale formation in the rings, treated cooling water is required. This water will be made available from off-site facilities.

It should be noted that the sorber operation shown on the flow sheet corresponds to the first set of conditions given in Table 38 which outlines the proposed operation schedule for each of the four sorbers. The material balance given in Table 39 shows the process flows for the major process streams, also for the first set of conditions in Table 38. To show the changes in stream flow which occur during a cooling cycle, Table 40 was provided as an alternate to Table 39. Table 40 shows only those flows which would be affected by the cooling cycle; all other flows are the same during both the heating and cooling cycles.

TABLE 38

SORBER OPERATIONS SCHEDULE

TIME (HRS)	SORBER D-1	 D-2	 D-3	 D-4
0	S	S	R	H
1	S	S	C	R
2	H	S	S	R
3	R	S	S	C
4	R	H	S	S
5	C	R	S	S
6	S	R	H	S
7	S	C	R	S
8	S	S	R	H

NOTES: S denotes a train which is sorbing SO_2
R denotes a train which is being regenerated
H denotes a train which is being heated
C denotes a train which is being cooled
0-7 represents one complete cycle; 8 is identical to 0 and is the start of another cycle

TABLE

HEATING CYCLE MAT

STREAM NO.		1 (INLET FLUE GAS FROM POWER PLANT)		1-A (FLUE GAS INLET TO SORBER D-1)		1
COMPOSITION	M.W.	LB/HR	MPH	LB/HR	MPH	L
CO	28.011					
CO_2	44.010	1,755,024	39,877.84	877,512	19,938.92	8
COS	60.075					
H_2	2.016					
H_2O	18.015	423,792	23,524.42	211,896	11,762.21	2
H_2S	34.080					
N_2	28.013	6,024,234	215,048.30	3,012,117	107,524.15	3,
O_2	31.999	304,411	9,513.14	152,206	4,756.57	
S_2	64.128					
SO_2	64.063	39,674	619.30	19,837	309.65	
CH_4	16.043					
TOTAL		8,547,135	288,583.00	4,273,568	144,291.50	4,

STREAM NO.		2 (SORBER EXIT HEADER)		2-A (FLUE GAS EXIT FROM SORBER D-1)		2
COMPOSITION	M.W.	LB/HR	MPH	LB/HR	MPH	L
CO	28.011					
CO_2	44.010	1,923,638	43,709.11	877,512	19,938.92	
COS	60.075					
H_2	2.016					
H_2O	18.015	483,814	26,856.18	211,896	11,762.21	
H_2S	34.080					
N_2	28.013	6,651,188	237,428.79	3,012,117	107,524.15	3,
O_2	31.999	322,185	10,068.60	147,251	4,601.75	
S_2	64.128					
SO_2	64.063					
TOTAL		9,380,825	318,062.68	4,248,776	143,827.03	4,

A

TABLE 39

HEATING CYCLE MATERIAL BALANCE

A (FLUE GAS INLET TO SORBER D-1)		1-B (FLUE GAS INLET TO SORBER D-2)		1-C (REGEN. GAS INLET TO SORBER D-3)		1-D (HEATING GAS INLET TO SORBER D-4)	
/HR	MPH	LB/HR	MPH	LB/HR	MPH	LB/HR	MPH
				9628.50	343.74		
877,512	19,938.92	877,512	19,938.92	17788.40	404.19	168,614	3831.27
				5338.07	2647.85		
211,896	11,762.21	211,896	11,762.21	1414.00	78.49	60,022	3331.76
012,117	107,524.15	3,012,117	107,524.15			626,953	22380.49
152,206	4,756.57	152,206	4,756.57			27,682	865.11
19,837	309.65	19,837	309.65				
				7.06	0.44		
273,568	144,291.50	4,273,568	144,291.50	34176.03	3474.71	883,271	30408.63

-A (FLUE GAS EXIT FROM SORBER D-1)		2-B (FLUE GAS EXIT FROM SORBER D-2)		2-C (REGEN. GAS EXIT FROM SORBER D-3)		2-D (HEATING GAS EXIT FROM SORBER D-4)	
/HR	MPH	LB/HR	MPH	LB/HR	MPH	LB/HR	MPH
				2492.42	88.98		
877,512	19,938.92	877,512	19,938.92	28527.28	648.20	168,614	3831.27
				645.81	10.75		
				1128.07	559.56		
211,896	11,762.21	211,896	11,762.21	29056.75	1612.92	60,022	3331.76
				18875.55	553.86		
012,117	107,524.15	3,012,117	107,524.15			626,953	22,380.49
147,251	4,601.75	147.251	4,601.75			27,682	865.11
				479.68	7.48		
				2545.22	39.73		
248,776	143,827.03	4,248,776	143,827.03	83757.84	3521.92	883,271	30,408.63

B

TABLE 3

STREAM NO.		3 (CLEAN FLUE GAS TO STACK)		4 (RECYCLE FLUE	
COMPOSITION	M.W.	LB/HR	MPH	LB/HR	MPH
CH_4	16.043				
C_2H_6	30.070				
C_3H_8	44.097				
C_4H_{10}	58.124				
C_5H_{12}	72.151				
C_6H_{14}	86.178				
CO	28.011				
CO_2	44.010	1,779,734	40,439.31	143,904	3269.
COS	60.075				
H_2	2.016				
H_2O	18.015	447,631	24,847.66	36,183	2008.
H_2S	34.080				
N_2	28.013	6,153,531	219,663.85	497,656	17764.
O_2	31.999	298,045	9,314.21	24,140	754.
S_2	64.128				
SO_2	64.063				
TOTAL		8,678,941	294,265.03	701,883	23797.

STREAM NO.		7 (RECYCLE FLUE GAS HEATER EFFLUENT)		8 (REGEN. GAS TO SORBER)	
COMPOSITION	M.W.	LB/HR	MPH	LB/HR	
CO	28.011			9628.50	343.
CO_2	44.010	168,614	3831.27	17788.40	404.
COS	60.075				
H_2	2.016			5338.07	2647.
H_2O	18.015	60,022	3331.76	1414.00	78.
H_2S	34.080				
N_2	28.013	626,953	22380.49		
O_2	31.999	27,682	865.11		
S_2	64.128				
SO_2	64.063				
CH_4	16.043			7.06	0.
TOTAL		883,271	30408.63	34176.03	34

A

TABLE 39 (CON'T)

(RECYCLE FLUE GAS)		5 (AIR TO RECYCLE FLUE GAS HEATER)		5-A (PILOT AIR TO RECYCLE FLUE GAS HEATER)		6 (NATURAL GAS TO RE-CYCLE FLUE GAS HEATER	
B/HR	MPH	LB/HR	MPH	LB/HR	MPH	LB/HR	MPH
						8221.39	512.46
						402.04	13.37
						101.42	2.30
						99.39	1.71
						77.20	1.07
						45.67	0.53
43,904	3269.80						
36,183	2008.52	3,734	207.27	415	23.04		
97,656	17764.94	116,286	4151.09	12,921	461.24	89.92	3.21
24,140	754.39	35,096	1096.78	3,899	121.85		
01,883	23797.65	155,116	5455.14	17,235	606.13	9037.03	534.65

8 (REGEN. GAS FEED TO SORBER)		9 (REGEN. GAS EFFLUENT FROM SORBER)	
LB/HR	MPH	LB/HR	MPH
9628.50	343.74	2492.42	88.98
17788.40	404.19	28527.28	648.20
		645.81	10.75
5338.07	2647.85	1128.07	559.56
1414.00	78.49	29056.75	1612.92
		18875.55	553.86
		479.68	7.48
		2545.22	39.73
7.06	0.44		
34176.03	3478.71	83757.84	3521.92

TABLE 40

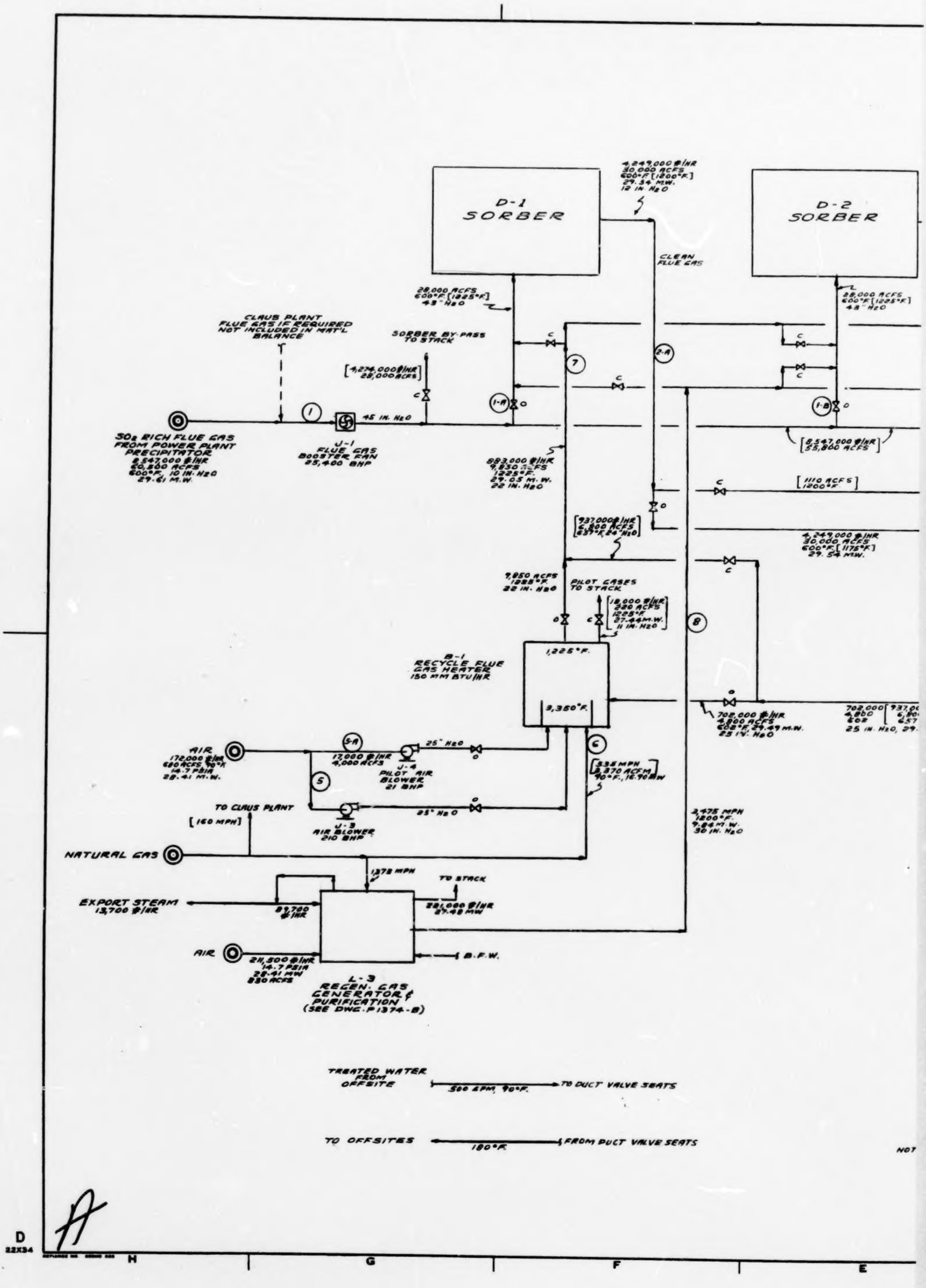
COOLING CYCLE MATERIAL BALANCE

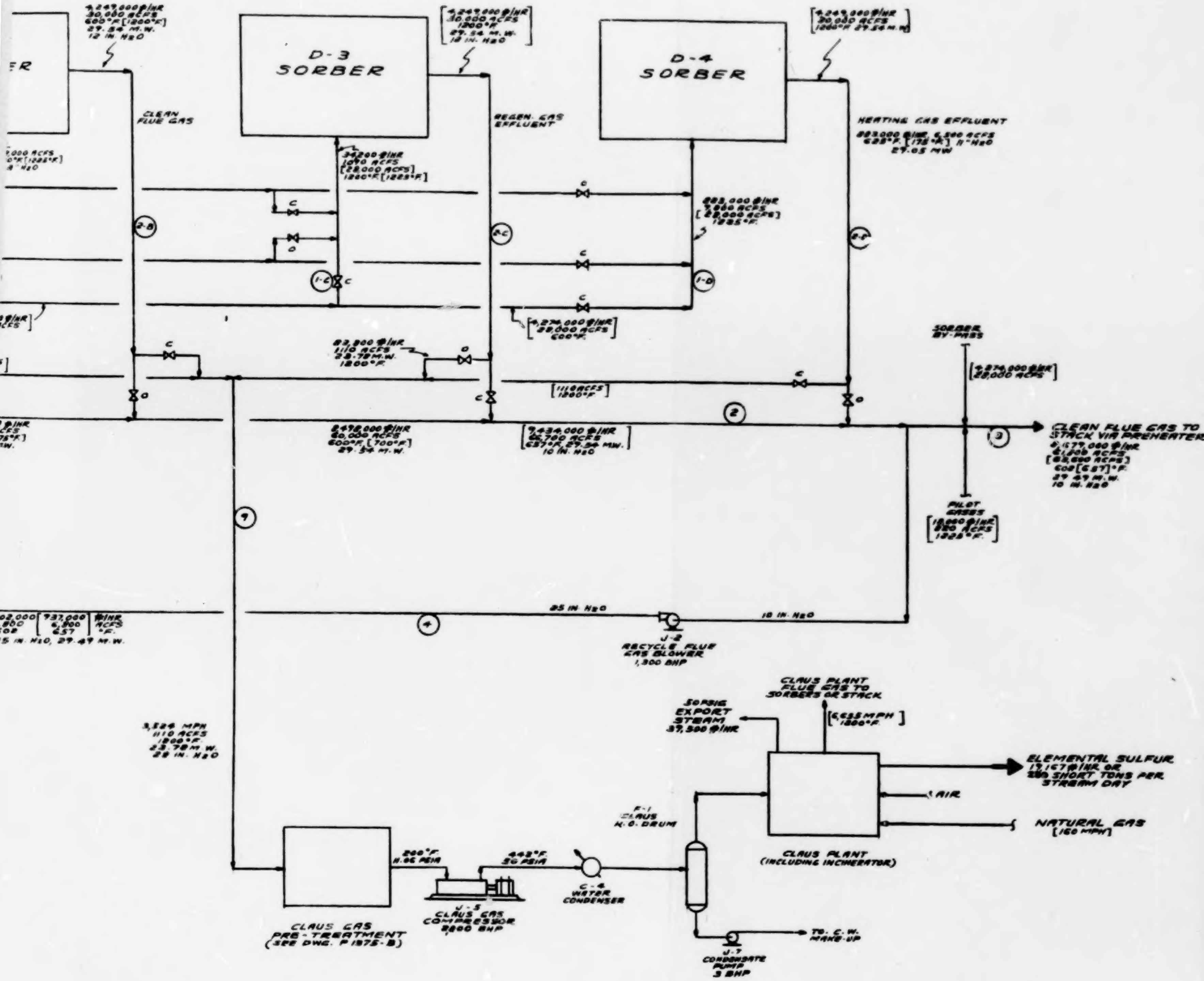
STREAM NO.		2 (SORBER EXIT HEADER)		3 (CLEAN FLUE GAS TO STACK)	
COMPOSITION	M.W.	LB/HR	MPH	LB/HR	MPH
CO_2	44.010	1,948,485	44,273.68	1,757,495	39933.99
H_2O	18.015	470,541	26,119.43	426,176	23656.74
N_2	28.013	6,688,372	238,756.17	6,037,164	215509.86
O_2	31.999	327,011	10,219.43	294,857	9214.56
TOTAL		9,434,409	319,368.71	8,515,692	288,315.15

STREAM NO.		4 (RECYCLE FLUE GAS)[1]		6 (NATURAL GAS TO PILOT BURNER OF B-1)	
COMPOSITION	M.W.	LB/HR	MPH	LB/HR	MPH
CH_4	16.043			822.14	51.25
C_2H_6	30.070			40.20	1.34
C_3H_8	44.097			10.14	0.23
C_4H_{10}	58.124			9.94	0.17
C_5H_{12}	72.151			7.72	0.11
C_6H_{14}	86.178			4.57	0.05
CO_2	44.010	193,461	4395.84		
H_2O	18.015	46,749	2595.01		
N_2	28.013	664,138	23,707.87	8.99	0.32
O_2	31.999	32,509	1,015.94		
TOTAL		936,857	31,714.66	903.70	53.47

NOTES:

1. The flow and composition given for Stream 4 are the same as that for Streams 1-D, 2-D and 7, which are to be changed from the values given in Table 39.
2. There is no Stream 5 flowing during the cooling cycle.


D-1
SORBER
D-2
SORBER
4,249,000 #/HR
30,000 ACFS
600°F [1200°F]
29.54 M.W.
12 IN. H2O
CLEAN
FLUE GAS
28,000 ACFS
600°F [1225°F]
48" H2O
28,000 ACFS
600°F [1225°F]
48" H2O
CLAUS PLANT
FLUE GAS IF REQUIRED
NOT INCLUDED IN MAT'L
BALANCE
SORBER BY-PASS
TO STACK
[4,274,000 #/HR
28,000 ACFS]
C
O
7
2-A
1-A
1-B
1
45 IN. H2O
SO2 RICH FLUE GAS
FROM POWER PLANT
PRECIPITATOR
8,547,000 #/HR
60,800 ACFS
600°F, 10 IN. H2O
29.61 M.W.
J-1
FLUE GAS
BOOSTER FAN
25,400 BHP
883,000 #/HR
9,850 ACFS
1225°F.
29.05 M.W.
22 IN. H2O
[8,547,000 #/HR
55,800 ACFS]
[1110 ACFS
1200°F]
[937,000 #/HR
6,800 ACFS
657°F, 24" H2O]
4,249,000 #/HR
30,000 ACFS
600°F [1175°F]
29.54 M.W.
9,850 ACFS
1225°F.
22 IN. H2O
PILOT GASES
TO STACK
[18,000 #/HR
220 ACFS
1225°F.
27.44 M.W.
11 IN. H2O]
8
B-1
RECYCLE FLUE
GAS HEATER
150 MM BTU/HR
1,225°F.
3,350°F.
702,000 #/HR
4,800 ACFS
602°F, 29.49 M.W.
25 IN. H2O
702,000 [937,0
4,800 6,80
602 657
25 IN. H2O, 29.
AIR
172,000 #/HR
680 ACFS, 90°F
14.7 PSIA
28.41 M.W.
5-A
17,000 #/HR
4,000 ACFS
25" H2O
J-4
PILOT AIR
BLOWER
21 BHP
5
J-3
AIR BLOWER
210 BHP
25" H2O
6
[335 MPH
3,370 ACFM
90°F., 16.90 MW]
TO CLAUS PLANT
[160 MPH]
NATURAL GAS
1378 MPH
3,475 MPH
1200°F.
9.84 M.W.
30 IN. H2O
TO STACK
EXPORT STEAM
13,700 #/HR
89,700 #/HR
281,000 #/HR
27.48 MW
AIR
211,500 #/HR
14.7 PSIA
28.41 MW
830 ACFS
B.F.W.
L-3
REGEN. GAS
GENERATOR &
PURIFICATION
(SEE DWG. P1374-B)
TREATED WATER
FROM
OFFSITE
500 GPM, 90°F.
TO DUCT VALVE SEATS
TO OFFSITES
180°F.
FROM DUCT VALVE SEATS
NOT
A
D
22X34
H
G
F
E


4,249,000 #/HR 30,000 ACFS 600°F [1200°F] 27.54 M.W. 12 IN. H2O
ER
CLEAN FLUE GAS
D-3 SORBER
[4,249,000 #/HR 30,000 ACFS 1200°F 27.54 M.W. 12 IN. H2O]
REGEN. GAS EFFLUENT
34,200 #/HR 1070 ACFS [28,000 ACFS] 1200°F [1225°F]
D-4 SORBER
[4,249,000 #/HR 30,000 ACFS 1200°F 27.54 M.W.]
HEATING GAS EFFLUENT
283,000 #/HR, 6,390 ACFS 625°F [175°F] 11" H2O 27.05 MW
283,000 #/HR 7,860 ACFS [28,000 ACFS] 1225°F
[4,274,000 #/HR 28,000 ACFS 600°F]
2-B
2-C
2-F
1-C
1-D
C
O
83,800 #/HR 1110 ACFS 23.78 M.W. 1200°F
[1110 ACFS 1200°F]
8,498,000 #/HR 60,000 ACFS 600°F [700°F] 27.54 M.W.
[9,434,000 #/HR 66,700 ACFS 637°F, 27.54 MW. 10 IN. H2O]
2
SORBER BY-PASS
[4,274,000 #/HR 28,000 ACFS]
3
CLEAN FLUE GAS TO STACK VIA PREHEATER
9,177,000 #/HR 61,200 ACFS [63,600 ACFS] 608 [637]°F 27.47 M.W. 10 IN. H2O
PILOT GASES
[18,000 #/HR 820 ACFS 1225°F]
9
4
25 IN. H2O
10 IN. H2O
J-2 RECYCLE FLUE GAS BLOWER 1,300 BHP
02,000 | 737,000 #/HR
800 | 6,800 ACFS
08 | 657 °F
5 IN. H2O, 27.47 M.W.
3,524 MPH 1110 ACFS 1200°F 23.78 M.W. 28 IN. H2O
CLAUS PLANT FLUE GAS TO SORBERS OR STACK
[6,655 MPH 1200°F]
50 PSIG EXPORT STEAM 37,500 #/HR
ELEMENTAL SULFUR 17,167 #/HR OR 206 SHORT TONS PER STREAM DAY
AIR
NATURAL GAS [160 MPH]
F-1 CLAUS K.O. DRUM
CLAUS PLANT (INCLUDING INCINERATOR)
200°F 14.65 PSIA
442°F 50 PSIA
CLAUS GAS PRE-TREATMENT (SEE DWG. P 1975-B)
J-5 CLAUS GAS COMPRESSOR 2800 BHP
C-4 WATER CONDENSER
TO C.W. MAKE-UP
J-7 CONDENSATE PUMP 3 BHP

NOTES: 1. [DESIGN VALUE] WHERE DIFFERENT FROM NORMAL.

NO.	DESCRIPTION	DATE	BY	APPROVED
1	GENERAL	2-27-70	A.G.S.	
2	INCREASED SULFUR PRODUCTION	8-31-70	A.G.S.	

REVISIONS

ENGINEERED: A.G.S.
DRAWN: F. R.
RELEASED FOR DESIGN:
DATED: 8-31-69
APPROVED:
ISSUED FOR CONSTRUCTION

THE M. W. KELLOGG COMPANY
a division of PULLMAN INCORPORATED

ALKALIZED ALUMINA PROCESS

FIXED BED SCHEME FOR
1,000 MW POWER PLANT FLUE GASES

CLASS & CD | AREA | JOB NUMBER | DRAWING NUMBER
6045 | P3196-D

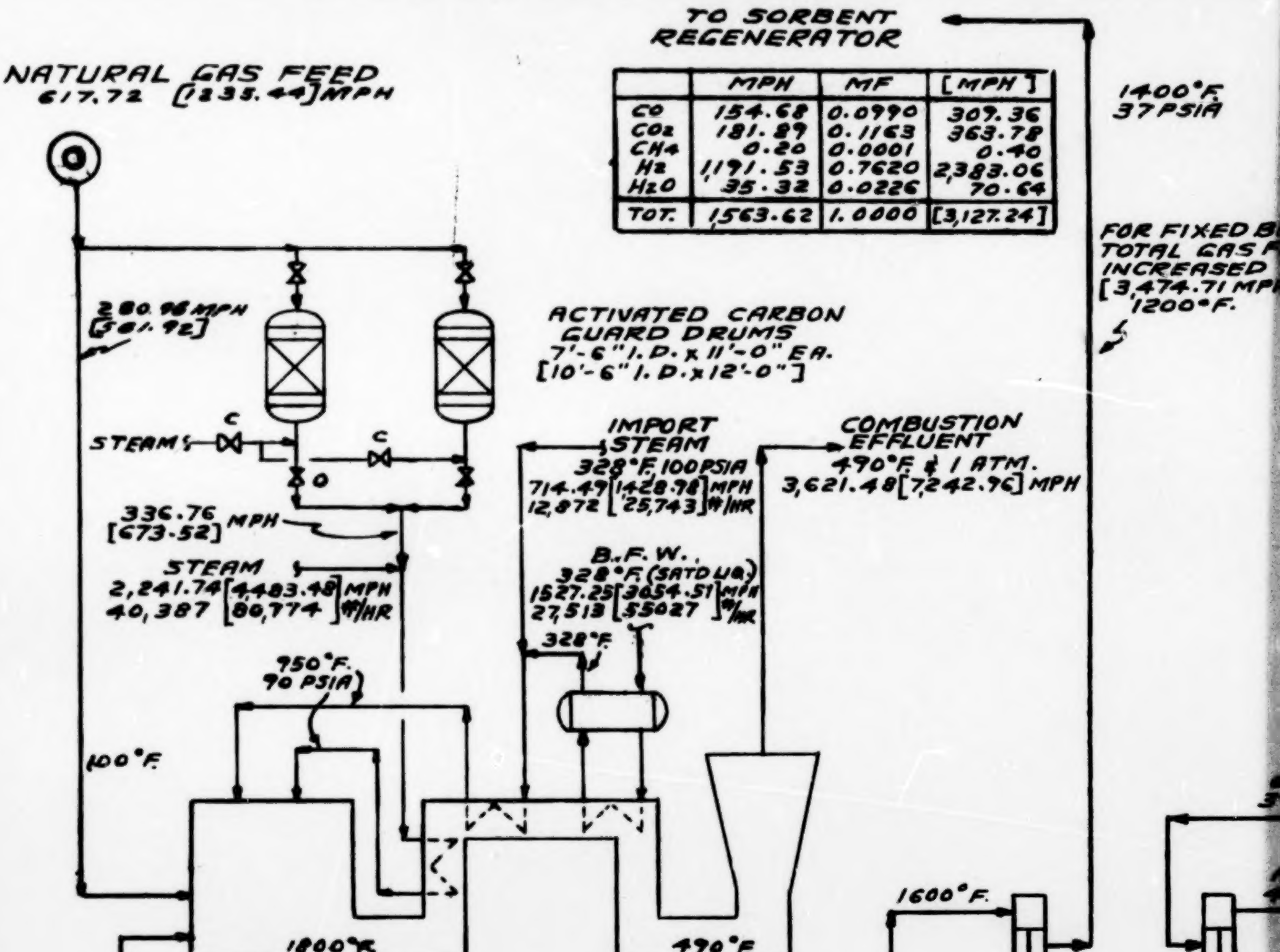
E D C B A

DRAWING NO. [redacted]

B

(1) Regeneration Gas Plant

For the base case designs it was decided to use a steam-reformed natural gas as the reducing gas needed to regenerate the loaded sorbent. A flow sheet P1374-B, of the reformer plant is shown in Figure 57. Natural gas feed is desulfurized in the activated carbon guard drums before entering the reformed furnace since sulfur will poison the reformer catalyst and therefore must be removed from the feed gas upstream of the reformer. In the reformer furnace natural gas and steam flow through catalyst packed tubes and react to form H_2, CO, and CO_2. The heat of reaction is supplied by burning additional natural gas with air outside the tubes. Reaction conditions to produce the desired gas composition for the present design were determined to be 50 psia and 1600°F. The effluent flue gas from the furnace is at 490°F and this gas is added to the power plant flue gas stream downstream of the sorber, and thus passes through the air preheater before being vented to the atmosphere.


TO SORBENT REGENERATOR
NATURAL GAS FEED
617.72 [1235.44] MPH
1400°F.
37 PSIA
FOR FIXED B
TOTAL GAS F
INCREASED
[3,474.71 MP
1200°F.
280.96 MPH
[561.92]
ACTIVATED CARBON
GUARD DRUMS
7'-6" I.D. x 11'-0" EA.
[10'-6" I.D. x 12'-0"]
STEAM
C
C
O
IMPORT
STEAM
328°F. 100 PSIA
714.49 [1428.98] MPH
12,872 [25,743] #/HR
COMBUSTION
EFFLUENT
490°F. & 1 ATM.
3,621.48 [7,242.96] MPH
336.76
[673.52] MPH
STEAM
2,241.74 [4483.48] MPH
40,387 [80,774] #/HR
B.F.W.
328°F. (SATD LIQ.)
1527.25 [3054.51] MPH
27,513 [55027] #/HR
328°F.
950°F.
90 PSIA
100°F.
1800°F.
490°F.
1600°F.
1,600°F.
50 PSIA
AIR
90°F. & 80% R.H.
3332.48 [6,664.97] MPH
REFORMER FURNACE
(RADIANT) PROCESS DUTY = 48.10 [96.19] MM BTU/HR
REGEN. GAS HEATER
8.86 MM BTU/HR (900 FT.
[17.72 MM BTU/HR 1800 FT

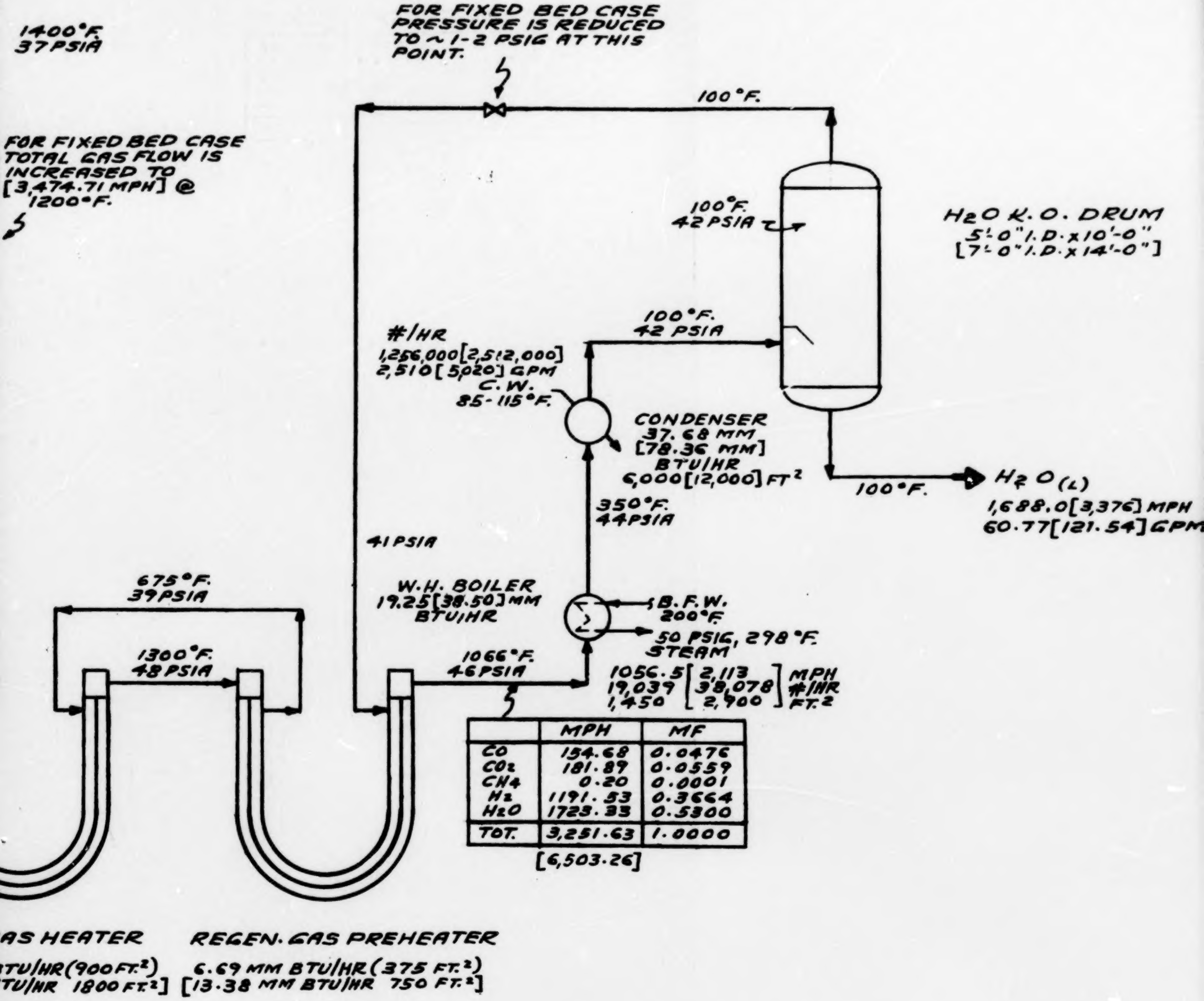
	MPH	MF	[MPH]
CO	154.68	0.0990	309.36
CO_2	181.89	0.1163	363.78
CH_4	0.20	0.0001	0.40
H_2	1191.53	0.7620	2,383.06
H_2O	35.32	0.0226	70.64
TOT.	1563.62	1.0000	[3,127.24]

NOTES: 1. FLOWS CORRESPOND TO THOSE ON THE FLUID BED SOPTION PROCESS FLOWSHEET P3198-D.

2. THE NUMBERS SHOWN IN BRACKETS INDICATE THE REQUIRED FLOWS, DUTIES AND SIZES FOR A SINGLE REGEN. GAS PLANT FEEDING TWO TRAINS OF EQUIPMENT SHOWN ON FLOWSHEET P3198-D.

B
11 X 17

DEFIANCE


1400°F.
37 PSIA
FOR FIXED BED CASE
TOTAL GAS FLOW IS
INCREASED TO
[3,474.71 MPH] @
1200°F.
FOR FIXED BED CASE
PRESSURE IS REDUCED
TO ~ 1-2 PSIG AT THIS
POINT.
100°F.
100°F.
42 PSIA
H2O K.O. DRUM
5'-0" I.D. x 10'-0"
[7'-0" I.D. x 14'-0"]
100°F.
42 PSIA
#/HR
1,256,000 [2,512,000]
2,510 [5,020] GPM
C.W.
85-115°F.
CONDENSER
37.68 MM
[78.36 MM]
BTU/HR
6,000 [12,000] FT²
100°F.
H2O (L)
1,688.0 [3,376] MPH
60.77 [121.54] GPM
350°F.
44 PSIA
41 PSIA
W.H. BOILER
19.25 [38.50] MM
BTU/HR
B.F.W.
200°F.
50 PSIG, 298°F.
STEAM
675°F.
39 PSIA
1300°F.
48 PSIA
1066°F.
46 PSIA
1056.5 [2,113] MPH
19,039 [38,078] #/HR
1,450 [2,900] FT.²
AS HEATER
TU/HR (900 FT.²)
TU/HR 1800 FT.²]
REGEN. GAS PREHEATER
6.69 MM BTU/HR (375 FT.²)
[13.38 MM BTU/HR 750 FT.²]

	MPH	MF
CO	154.68	0.0476
CO_2	181.89	0.0559
CH_4	0.20	0.0001
H_2	1191.53	0.3664
H_2O	1723.33	0.5300
TOT.	3,251.63	1.0000

[6,503.26]

REGENERATION GAS PLANT
FOR AA FLOWSHEETS

NO.	REVISION DESCRIPTION	DATE	BY	CHK.
1	GENERAL	2-27-70	GMD	

ISSUED FOR FABRICATION
ISSUED FOR CONSTRUCTION

ENGR. G.M.D.
DRAWN: F.P.R.
CHECKED:
APPROVED:
DATED: 8-31-69

THE M. W. KELLOGG COMPA
a division of PULLMAN INCORPORATED

FIGURE 57

CLASS	AREA	JOB NO.	DRAWING
		6045	P1374-

[redacted]

The product gases from the reformer contain over 50% water and since a "dry" gas is needed for regeneration, this water must be condensed. The regenerator temperature has been set at 1400°F which is also the desired temperature for the reducing gas feed. Thus the heat exchanger train has been arranged such that the cooled, dried reformed gas is preheated by exchange with the hot effluent from the reformer furnace. The preheating operation requires two stages of exchange in series: 1) a regeneration gas heater where the dry reformed gas is heated from 675° to 1400°F by cooling the reformer effluent from 1600° to 1300°F, 2) a regeneration gas preheater where the dry gas is heated from 100° to 675°F by further cooling the reformer outlet gas from 1300 to 1066°F. The temperature of the wet gas stream effluent from the regeneration gas preheater is reduced from 1066° to 350°F by generating 50 psig steam in the waste heat boiler. Finally, the water is condensed by exchange with cooling water and the resulting gas-liquid mixture separated in the water knock-out drum. The dry gas, containing some 2% water, is preheated as described above and sent to the regenerator. The dried, preheated reformed gas is at about 37 psia which provides an ample allowance for friction losses and control valves since the maximum pressure in the sorber is 30 psia.

The flow rates shown on P1374-B are for the fluid bed and dispersed phase cases. The reducing gas quantity required for the fixed bed cas is shown on the outlet stream only but if intermediate values are desired they can be prorated using the ratio of the two product streams shown.

To date, the maximum allowable water content in the reducing gas to the regenerator has not been determined and the present designs are based on "dry" gas (i.e., about 2% water). If later data show that higher water concentrations can be tolerated, then the reformer can be redesigned to produce the regenerator feed gas directly thus eliminating the cooling and reheating operations.

An evaluation of this alternative reformer design will be made in the phase 3 work.

(2) Claus Gas Pre-Treatment

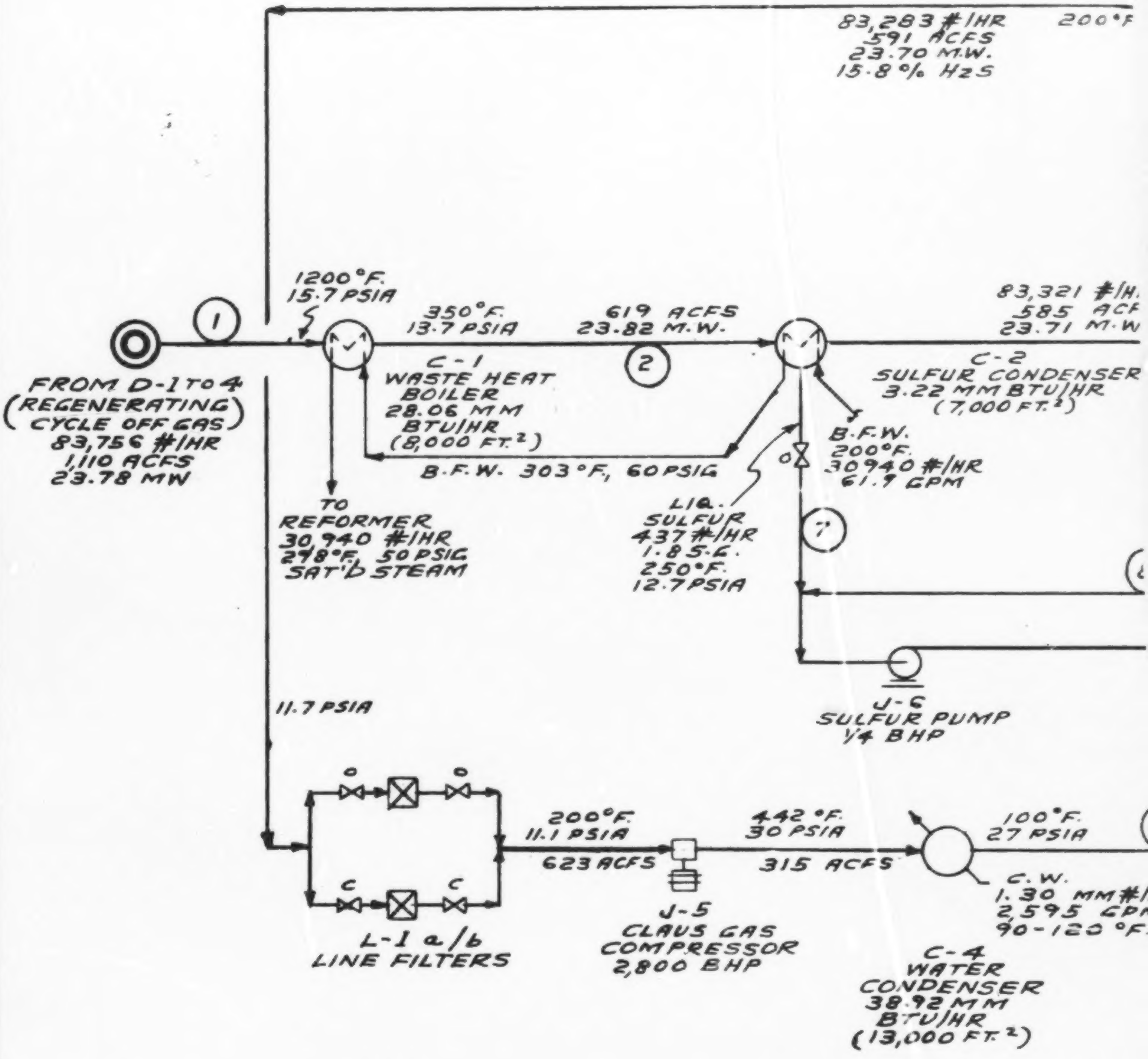
The regenerator gas effluent in all three of the present process designs contains about 18% H_2S, and greater than 40% H_2O. In addition, equilibrium calculations have shown that there will probably be some elemental sulfur present as vapor. It was originally planned to condense the water from the regenerator exit gas, thereby increasing the H_2S concentration to about 30% and thus reduce the size of the Claus plant. Further study showed, however, that because of the sulfur vapor present in the gas, cooling and condensing the water would also condense the sulfur. Further, because of the temperature range and sulfur concentrations involved, the equipment required became complicated and expensive; i.e., the sulfur temperature-viscosity relationship necessitated several steps involving quench towers, scraped surface exchangers (or equivalent type equipment), the cost of which would more than offset the incremental cost of the Claus plant due to a dilute H_2S feed. Therefore, for the fluid bed and dispersed phase cases, no pretreatment to the regenerator off gas is used other than a waste heat boiler for heat economy. The feed to the Claus reactor should be about 400°F and since the regenerator effluent is at 1400°F, the gas is cooled to the desired temperature by generating 50 psig steam. A credit will be taken for any steam in excess of that needed for the alkalized alumina plant.

The fixed bed case does require a pretreatment section, however, because the fixed bed regenerator is at atmospheric pressure and the off-gas must be compressed since the pressure drop through the Claus plant is about 10 psi. Limitations on the discharge temperatures of the compressor results in a maximum inlet gas temperature of about 250°F for the pressure range involved, and cooling the gas to this temperature condenses the sulfur vapor present. A process flow scheme, P1375-B, for this fixed bed regenerator off gas treatment and compression is shown in Figure 58.

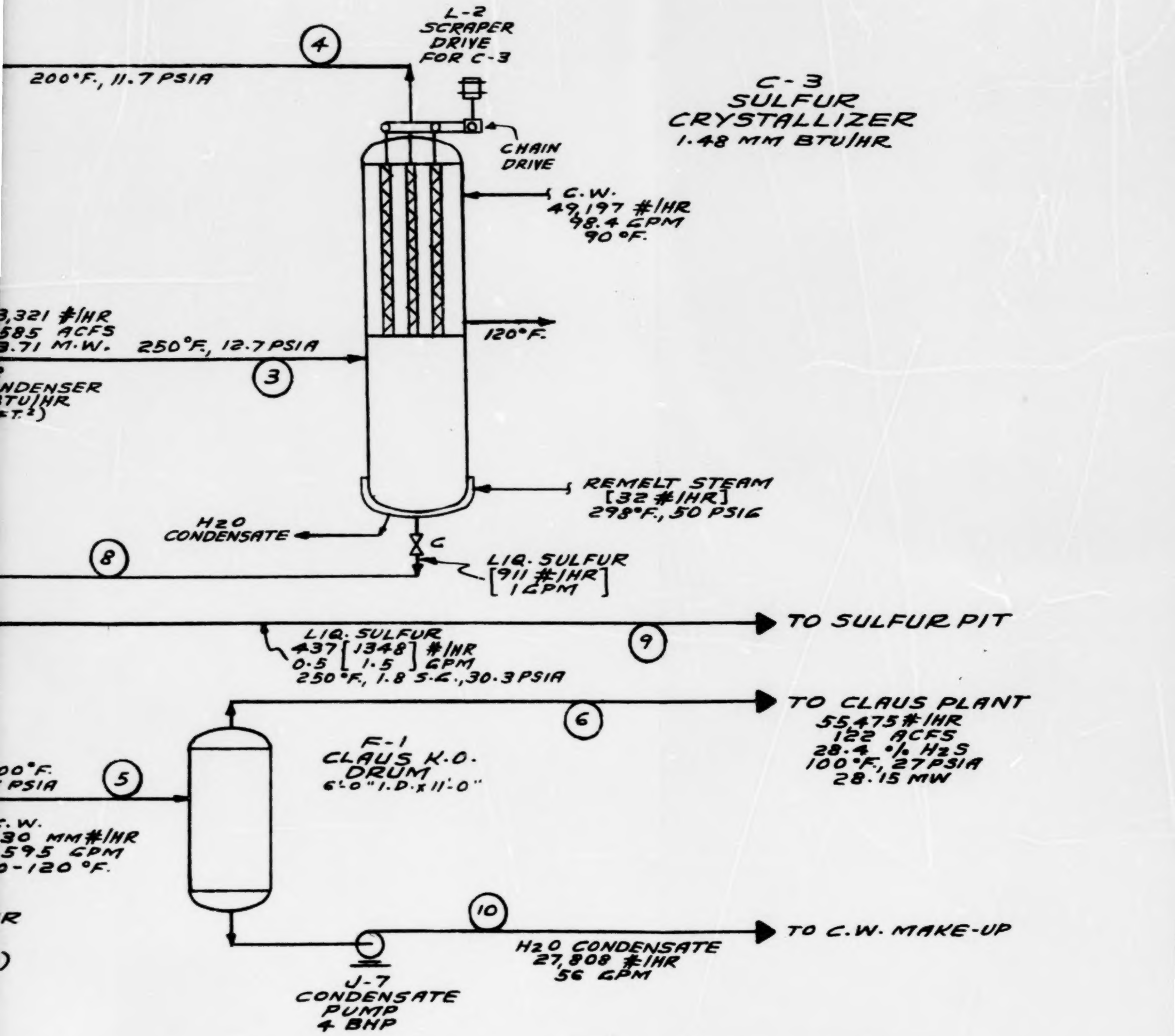
A brief process description for the fixed-bed Claus gas pretreatment section is as follows: referring to Figure 58, the hot gaseous effluent from the regenerator passes through a waste heat boiler, C-1, where it is cooled to 350°F by generating 65 psia steam. From C-1, the gas flows through a sulfur condenser, C-2, where most of the sulfur vapor is condensed to liquid sulfur at 250°F by exchange with the boiler feed water for C-1. At 250°F liquid sulfur will flow readily and is pumped to the main sulfur storage pit. The effluent gas from C-2 is sent to the water-cooled, scraped surface C-3 sulfur crystallizer (or desublimer) and cooled to 200°F thereby causing the remaining sulfur vapor to be deposited as sulfur crystals on the cold tubes. The sulfur crystals are continuously scraped from the tube walls and fall by gravity into the steam-heated bottom section of C-3 where they are melted and heated to 250°F and then pumped along with the condensed sulfur from C-2 to the sulfur storage pit.

The effluent gas from C-3 passes through a guard filter, L-1, which removes any entrained sulfur particles and is then compressed to 30 psia in the Claus gas compressor, J-5. The J-5 discharge stream is cooled to 100°F in the C-4 water condenser wherein most of the water is condensed, thereby increasing the H_2S concentration in the gas stream to about 27%. The liquid water and gas are separated in F-1 knockout drum and the dried compressed gas is sent to the Claus unit for conversion to elemental sulfur.

It should be noted that since it is necessary to remove the sulfur, the water can be easily condensed thus upgrading the Claus feed gas. The fixed bed case, therefore, has a more concentrated feed gas for the Claus plant than the fluid bed or dilute phase processes.


83,283 #/HR
591 ACFS
23.70 M.W.
15.8 % H2S
200°F
1200°F.
15.7 PSIA
1
FROM D-1 TO 4
(REGENERATING CYCLE OFF GAS)
83,756 #/HR
1,110 ACFS
23.78 MW
350°F.
13.7 PSIA
619 ACFS
23.82 M.W.
2
C-1
WASTE HEAT BOILER
28.06 MM BTU/HR
(8,000 FT.²)
B.F.W. 303°F, 60 PSIG
TO REFORMER
30,940 #/HR
298°F, 50 PSIG
SAT'D STEAM
LIQ. SULFUR
437 #/HR
1.8 S.G.
250°F.
12.7 PSIA
83,321 #/H
585 ACF
23.71 M.W
C-2
SULFUR CONDENSER
3.22 MM BTU/HR
(7,000 FT.²)
B.F.W.
200°F.
30940 #/HR
61.9 GPM
7
J-6
SULFUR PUMP
1/4 BHP
11.7 PSIA
O
C
L-1 a/b
LINE FILTERS
200°F.
11.1 PSIA
623 ACFS
442 °F.
30 PSIA
315 ACFS
J-5
CLAUS GAS COMPRESSOR
2,800 BHP
100°F.
27 PSIA
C.W.
1.30 MM #/
2,595 GP
90-120 °F
C-4
WATER CONDENSER
38.92 MM BTU/HR
(13,000 FT.²)

NOTES: 1. [DESIGN VALUE] WHERE DIFFERENT FROM NORMAL.
2. REMELT STEAM TO C-3 IS ON INTERMITTENT FLOW.


L-2
SCRAPER
DRIVE
FOR C-3
④
200°F., 11.7 PSIA
C-3
SULFUR
CRYSTALLIZER
1.48 MM BTU/HR
CHAIN
DRIVE
C.W.
49,197 #/HR
98.4 GPM
90 °F.
3,321 #/HR
585 ACFS
3.71 M.W.
250°F., 12.7 PSIA
③
NDENSER
TU/HR
120°F.
REMELT STEAM
[32 #/HR]
298°F., 50 PSIG
H2O
CONDENSATE
⑧
C
LIQ. SULFUR
[911 #/HR]
[1 GPM]
TO SULFUR PIT
LIQ. SULFUR
437 [1348] #/HR
0.5 [1.5] GPM
250°F., 1.8 S.G., 30.3 PSIA
⑨
TO CLAUS PLANT
55,475 #/HR
122 ACFS
28.4 % H2S
100°F., 27 PSIA
28.15 MW
⑥
F-1
CLAUS K.O.
DRUM
6'-0" I.D. x 11'-0"
00°F.
PSIA
⑤
C.W.
30 MM #/HR
595 GPM
-120 °F.
R
⑩
TO C.W. MAKE-UP
H2O CONDENSATE
27,808 #/HR
56 GPM
J-7
CONDENSATE
PUMP
4 BHP

FIXED BED FLOW SHEET
CLAUS GAS PRE-TREATMENT

NO.	REVISION DESCRIPTION	DATE	BY	CHK.
1	GENERAL	2-27-70	GMD	

ISSUED FOR FABRICATION
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ENGR: G.M.D.
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APPROVED:
DATED: 8-31-69

THE M. W. KELLOGG COMPA
a division of PULLMAN INCORPORATED

FIGURE 58

CLASS	AREA	JOB NO.	DRAWING
		6045	PJ375-

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B

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